"Energy from waste as a renewable energy supply to supplement electricity in South Africa"

by

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Declaration

I know the meaning of plagiarism and declare that all of the work in the document, save for that which is properly acknowledged, is my own.

In memorium

Marquind Jorge Jack - beloved cousin and friend

1979 - 2009

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Synopsis

The need for renewable energy in South Africa's energy mix, particularly in the coal-dominated electricity generation sector, is well recognised. Growing awareness of the negative impacts of first generation biofuels is driving a move to second generation biofuels to avoid competition with food crops and turning of virgin lands. Biogenic waste fuels, such as sewerage sludge, agricultural, forestry and sugar mill residues and organically loaded wastewater streams, represent potential feedstocks for the generation of heat and renewable electricity through combustion, gasification and anaerobic digestion. In this study, their potential is investigated through addressing of the following objectives: identification of technologies suitable for processing biogenic waste in South Africa; development of energy yield equations to compare these technologies; validation of these equations using industrial data; characterisation of biogenic waste fuels and their effect on processing; generation of a decision-making tool for technology selection based on feedstock characteristics; and, lastly, determination of the potential energy from biogenic waste streams in South Africa.

Combustion, gasfication and anaerobic digestion were evaluated as technologies to generate heat and power from distributed waste biogenic feedstocks in South Africa. Three main criteria were considered in the final recommendation viz., the thermal efficiency with small scale operation; technology challenges and the operational experience within South Africa. Gasification (in its current form) was dismissed due to the technical challenges of tar removal which limit the range of application of the fuel gas (Kiel *et al.*, 2004)) and the limitation of higher efficiencies to scales greater than 100 MW (Bridgewater, 1995; Wang, *et al.*, 2008). The use of anaerobic digestion as an energy generating technology, not solely for wastewater treatment, requires changes in plant control philosophy. For feedstocks resistant to biological action, pretreatment is required. The clean biogas product can be used directly in reciprocating gas engines. Combustion is relatively simple and achieves thermal efficiencies of 20 to 40%. Some operational experience of processing biogenic waste fuels exists within South Africa.

The application of combustion and anaerobic digestion to processing biogenic waste fuels, specifically sewerage sludge, are compared through a case study of three Thames Water wastewater treatment plants. The study showed that anaerobic digestion with thermal

hydrolysis pretreatment gave the highest energy yield. This anaerobic digestion process achieved a volatile solids destruction of 65%, compared to 45% in the absence of feed pretreatment. Combustion of a sludge containing 20% dry solids and anaerobic digestion of sludge to achieve 45% volatile solids destruction resulted in comparable energy recovery. Plant data suggested that combustion was the inferior option, owing to sensitivity of gross energy recovery on dry solids content.

Recalcitrant material reduces the maximum attainable volatile solids destruction and concomitant gas yield on anaerobic digestion. High-temperature, high pressure (HPHT) pretreatments cause structural changes increasing amenability to hydrolysis and digestion. High pressure homogenisation breaks recalcitrant cell walls to release the digestible cell content. The energy requirements of these pretreatments were compared with the energy yield from anaerobic digestion. With HTHP treatments, a positive nett energy yield is predicted at dry solids contents above 20% and a volatile solids destruction above 50%, assuming no heat recovery and a thermal efficency of 40%. With high pressure homogenisation at pressures greater than 5 MPa, a positive energy yield is not predicted.

The high volatile content of biogenic waste fuels results in similarities to gas combustion in a bubbling fluidised bed. Higher bed temperatures promote inbed combustion and higher combustion efficiencies. Underbed feeding allows longer residence time in the bed and hence higher conversion of volatiles and tar than overbed feeding. Experimental investigations of the combustion efficiency of woodchip in a bubbling fluidised bed showed little effect of temperature above ~700°C. The combustion efficiency of sewerage sludge below 750°C was markedly lower. Combustion efficiency of woodchip decreased from 90% to 65% when the feed location was changed from underbed feeding to overbed feeding. In this study the contribution to loss in combustion efficiency from overbed burning was small (less than 10%). It is proposed that a greater loss arose from volatiles or tars passing through the system unburnt. While the absolute contribution of overbed burning to total energy release is small, changes in freeboard temperature of up to 50°C were seen, owing to the small thermal mass of freeboard air compared to the bed.

A methodology to compare the energy yield from combustion and anaerobic digestion based on feedstock characteristics is presented. Following selection of appropriate technology,

energy yield models were used to determine the energy recovery. Competing uses for the residue stream have been highlighted in the study, *e.g.* nutrient recycling in the case of agricultural residues, cane trash and forestry residues. Maximum and probable energy recoveries, *viz.* 310 and 200 PJ/yr are are similar to those obtained by Williams and Eberhard (1988) and the DME (2004), and to Lynd *et al*. (2003) when the contribution from energy crops is neglected. Assuming a thermal efficiency of 30%, the maximum and probable electrical output from biogenic waste represents 11% and 7.5% of current demand. It is thus concluded sourcing 7.5% of current electricity demand from renewable sources of available biogenic waste is a realistic target.

To summarise, considerable potential energy recovery from biogenic waste fuels exists, both in South Africa and elsewhere. While potential technologies exist, the technical challenges resulting from specific characteristics of biogenic waste fuels (eg. recalcitrant matter and high volatile content) must be recognised. Further, drivers for renewable energy need to address non-technical issues such as the competing use of the fuel, installation of technology and infrastructure development.

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Glossary

Archae – the class of micro-organisms, distinctly different from bacteria, to which, *inter alia,* methanogens belong

Biofuel – specifically, any renewable fuel derived from biological matter but used colloquially to refer to liquid renewable fuel

Biogenic – derived from biological (as opposed to fossil) sources

Biomass – lignocellulosic plant material *i.e.* wood, agricultural residues, energy crops

Chemical Oxygen Demand (COD) – measures the organic fraction of a wastewater sample. A COD test determines the amount of oxygen required to oxidise the organic fraction in the sample

Co-generation – a power plant system where the hot flue gases from the gas turbine generator set are used to raise steam to drive a steam turbine. High thermal efficiencies are achieved in this system

Colony Forming Units (CFU) – a measure of the number of viable bacteria present in a sample

Combined Heat and Power (CHP) – the concomitant generation of electricity and high grade process heat

Dewater – removal of free water in sludge, usually by mechanical means. Typically used in the wastewater treatment industry.

Energy efficiency – the ratio of actual amount of energy produced to the theoretical amount o of energy produced

Energy recovery – used loosely as the amount of energy gained from a process

Energy yield – the ratio of the amount of energy produced by a process to that required by the process

Equivalence ratio – the ratio of the mols of oxygen fed to a combustor or gasifier to the number of mols of oxygen required for complete combustion

Filter Paper Units (FPU) – a measure of cellulase enzyme activity as determined by the Filter Paper Assay. Filter paper is digested by the enzyme and the amount of glucose product is measured. $FPU = \frac{0.37}{[Enzyme]}$ where [Enzyme] is the enzyme concentration required to release 2 mg glucose.

First generation biofuels – biofuels which are produced from a single storage product of the plant e.g. seed oil or plant sugar

Generation time – the time it takes for a bacterial population to double in size

Gram-negative bacteria – bacteria with an outer and inner membrane structure which does not retain the the crystal violet dye applied in the Gram staining test

Greenhouse Gas (GHG) – gaseous emissions to atmosphere from anthropogenic and natural processes which reflect long wave terrestrial radiation from leaving the atmosphere thus causing a heating effect. Examples include CO_2, CH_4 and N_2O.

Higher heating value (HHV) – HHV measures the enthalpy change with the water vapour exiting as a liquid, and is the value measured by bomb calorimetry.

Hot gas efficiency – a measurement of gasifier efficiency calculated from the ratio of the energy content in the hot gas leaving the gasifier and energy content of the gasifier feed

Lower heating value (LHV) – the amount of heat released on complete combustion of a fuel where the combustion products are CO_2(g) and H_2O(g). This value is generally determined in a bomb calorimeter.

Mesophile – a micro-organism which thrives at temperatures up to 37°C

Million Ton Oil Equivalent – the amount of energy released from burning 1 ton of crude oil. The IEA defines this as 42 GJ.

Pathogen reduction – a reduction in a pathogenic indicator organism, typically *E Coli.* Sludge sterilisation processes often require a log 5 pathogen reduction. *i.e.* a reduction in *E Coli* count from 10^{10} to 10^5 organisms.

Scale factor – an engineering index used to relate plant capacity to capital cost based on the fact that processing plants realise economies of scale. Cost = $(\text{Capacity})^x$ where x is the scale factor.

Second generation renewables – biofuels which are produced from lignocellulosic material, algae or biogenic waste streams

Sonication – a laboratory method for disrupting the cell wall (generally to release proteins) which uses ultrasound induced cavitation to break the cell wall

Sustainable energy – an energy source which does not deplete non-renewable sources nor endanger the long term health of the earth. Renewable energy is generally regarded as a sustainable energy supply

Thermal efficiency – in furnace design thermal efficiency refers to the ratio of electrical energy gained from the system to the amount of chemical energy fed to the system

Thermophile – a micro-organism which thrives at temperatures higher than 37°C

Thicken – to reduce the moisture content (or increase the dry solids content) of a sludge, typically not using mechanical means. Typically used in the wastewater treatment industry.

Vector – an organism capable of spreading pathogens either by physically transporting the pathogens or by playing a positive role in the life cycle of the pathogen

Volatile solids – the organic fraction of a biogenic stream which can be degraded by microbial action. Typically used in the wastewater treatment industry

Nomenclature

Symbol	*Description*
%ash	ash content of fuel [%]
%C	carbon content of fuel [%]
$\%K_2O$	K_2O content of fuel [%]
$\%Na_2O$	Na_2O content in fuel [%]
%power	percentage power input from furnace [%]
[CO]	molar concentration of CO in the flue gas [%]
$[CO_2]$	molar concentration of CO_2 in the flue gas [%]
$[O_2]$	molar concentration of CO_2 in the flue gas [%]
A_h	heat transfer area [m^2]
A_{hr}	radiative heat transfer surface [m^2]
AI	alkali index [kg/GJ]
ash	ash content in fuel on dry basis [%]
Cp_{air}	heat capacity of gas [kJ/kg.K]
Cp_{bed}	heat capacity of bed material [J/kg.K]
Cp_{gas}	heat capacity of gas [J/kg.K]
Cp_{slurry}	heat capacity of slurry [J/kg.K]
DS	dry solids content [%]
E	energy consumption [J]
E_{AD}	energy yield from anaerobic digestion [MJ/day]
$\hat{E}_{AD}$	energy yield from anaerobic digestion [MJ/kg VS]
$E_{combustion}$	energy yield from combustion [MJ/day]
$\hat{E}_{combustion}$	energy yield from combustion [MJ/kg DS]
$\hat{E}_{combustion}$	energy yield from combustion [MJ/kg VS]
$\hat{E}_{HPH}$	energy required for high pressure homogenisation [W/kg]
$\hat{E}_{TH}$	energy required for thermal hydrolysis [W/kg]
h	overall furnace to bed heat transfer coefficient [$W/m^2.K$]
HHV	higher heating value of the fuel [GJ/kg]
HI	harvest index [#]
H_{loss}	harvest loss [#]
I	current through furnace heating rods [A]
k	conductivity of the firebricks [W/m.K]
l	thickness of firebricks [m]
LHV_{CH4}	lower heating value of CH_4 at STP [35.8 MJ/m^3]
$LHV_{fuel,db}$	lower heating value of fuel on dry basis [MJ/kg]
$LHV_{fuel,daf}$	lower heating value of fuel on dry ash-free basis [MJ/kg]
$LHV_{fuel,daf}$	lower heating value of fuel on dry ash-free basis [MJ/kg]
m_{air}	mass of gas flowrate through bed [kg/s]
M_{bed}	mass of bed material [kg]
$m_{biomass}$	mass flowrate of biomass [kg/s]
m_{DS}	mass flowrate of dry solids [kg/day]
M_{fuel}	mass of fuel placed in hopper [kg]
m_{gas}	mass flowrate of gas [kg/s]
MM_C	molar mass of carbon [kg/mol]
$M_{residue}$	amount of residue available for collection [ton]
M_{seed}	seed harvest [ton]

MW_e	electrical power output
N	number of passes [#]
$N_{C,in}$	carbon fed to the system [mol]
$N_{C,out}$	carbon exiting system in flue gas [mols]
n_{gas}	molar flow of gas through the reactor [mol/s]
$N_{O2,out}$	oxygen exiting system in flue gas [mols]
N_p	number of particles [#]
P	operating pressure of homogeniser [Pa]
Q	volumetric flowrate [m^3/s]
$Q_{experimental}$	experimentally observed heat release to the bed [W]
Q_{fb}	heat lost due to overbed burning [W]
$Q_{furnace}$	heat input to bed from furnace [W]
Q_{in}	heat input to bed from furnace [W]
Q_{loss}	heat loss from bed [W]
$Q_{maximum}$	maximum possible heat release to the bed [W]
$Q_{radiative}$	radiative heat input from furnace [W]
r_c	rate of combustion [kmol/particle/s]
T^*	$T_{bed} - T_{bed,ss}$ [K]
$T_{element}$	element temperature [K]
T_{bed}	bed temperature [K]
$T_{bed,ss}$	steady state bed temperature [K]
$T_{furnace}$	furnace temperature [K]
T_{inlet}	inlet air temperature [K]
$T_{wall, outer}$	outer wall temperature [K]
$T_{wall,inner}$	inner wall temperature [K]
U	superficial velocity [m/s]
U_h	overall heat transfer coefficient [$W/m^2.K$]
U_{mf}	minimum fluidisation velocity [m/s]
U_R	average rise velocity of coal particle in bed [m/s]
V	potential difference across heating rods [V]
VSd	percentage volatile solids destruction [%]
VSd	percentage volatile solids destruction [%]
Y_{CH4}	yield of methane STP [0.65 m^3/kg VS]
ΔH_c	heat of combustion [J/kmol]
ΔH_c	heat of combustion of fuel [J/kg]
$\Delta Hvap_{water}$	heat of evaporation of water at 25°C [2.38 MJ/kg]
ΔP	operating pressure of unit [Pa]
ΔT	temperature change [K]
ΔT_{fb}	jump in freeboard temperature [K]
ε	emissivity of heating rods
η	combustion efficiency
ρ_{slurry}	slurry density [kg/m^3]
σ	Stefan Boltzmann constant [$5.67 \cdot 10^{-8}$ $W/m^2/K^4$]

List of abbreviations

CDM – Carbon Development Mechanism

CFU – Colony Forming Units

CHP – Combined Heat and Power

COD – Chemical Oxygen Demand

CSTR – Continuous Stirred Tank Reactor

ER – Equivalence ratio

FPU – Filter Paper Units

GHG – Greenhouse Gas

HHV – Higher Heating Value

HRT – Hydraulic Retention Time

LHV – Lower Heating Value

MTOE – Million Ton Oil Equivalent

SRT – Solids Retention Time

UASB – Upflow Anaerobic Sludge Blanket

VS – Volatile solids

Chapter 1 Introduction

1.1. Biofuels as a sustainable energy source

The two tenants of a sustainable energy policy require energy efficiency and increased use of renewable energy (ACEEE, 2007). Renewable energy sources are classified into two groups: those which can be replenished in a short timeframe (e.g. biomass), and those which are inexhaustible in the timeframe of our civilisation (e.g. solar, wind, geothermal and tidal). In this thesis, the sustainable use of the former is considered in terms of availability and technology selection.

1.1.1. Definitions and terms

"Biofuel" is the term given to a renewable fuel derived from biological matter, encompassing a wide range of end products including biodiesel, bioethanol, biogas and dimethyl ether derived from biomass. Colloquially it refers primarily to the former two fuels. Biofuels may be further classified as either first or second generation biofuels. First generation biofuels are those derived from the storage products of the plant only, *i.e.* seed oils (e.g. soya and canola biodiesel) or plant sugars (e.g. ethanol from maize or sugarcane), hence much of the plant matter is not converted to fuel. Second generation biofuels do not use conventional agricultural crops as feedstock, deriving fuels from lignocellulosic material, algae and biogenic waste. Essentially, it is a biofuel processed from the entire plant or biomass as opposed to a single storage product, or it can be viewed as a biofuel which does not compete with food crops.

1.1.2. Retreat from first generation biofuels in first world countries

Statutory targets for biofuel use were introduced in many first world countries around the turn of the century. The goal of the legislation was to reduce greenhouse gas emissions. However, the introduction of these targets has had a number of negative spin-offs, viz., environment destruction, the use of biofuels with marginal greenhouse gas emission savings and a potential contribution to the food price increases

1.1.2.1. Increased greenhouse gas emission from land use change

Life Cycle Analysis (LCA) has been used to determine the saving in CO_2 emissions users on switching from fossil fuel to biofuels. Kaltschmitt *et al.* (1994) determined a CO_2 equivalent

emission saving of 58% for Rape Methyl Ester (RME) compared to mineral diesel produced in Germany. Biodiesel production in the UK realised a CO_2 equivalent saving of 26 to 32% depending on small or large scale production, with small scale production realising the greater saving. (Stephenson *et al.*, 2008). The difference in CO_2 equivalent savings is due to Stephenson *et al.* using current IPCC guidelines for N_2O soil emissions, which have been revised upwards since the Kaltschmitt study.

There has been increasing importance placed on where biofuel crops are grown and the natural habitats destroyed to accommodate them. Virgin, uncultivated lands store large reservoirs of terrestrial carbon. New land preparation, through burning of the biomass and ploughing, releases more CO_2 than the crop can mitigate in a short timeframe. To analyse this, an indicator has been developed called the carbon debt repayment time and defined as the time a biofuel crop has to be grown to "recoup" the carbon emission released on initial land preparation. The carbon debt repayment time estimates are shown in Figure 1-1 for some biodiesel producing areas. This highlights the lengthy carbon debt repayment times for traditional first generation biofuels compared to the second generation cellulosic ethanol option, viz. Prairie biomass ethanol.

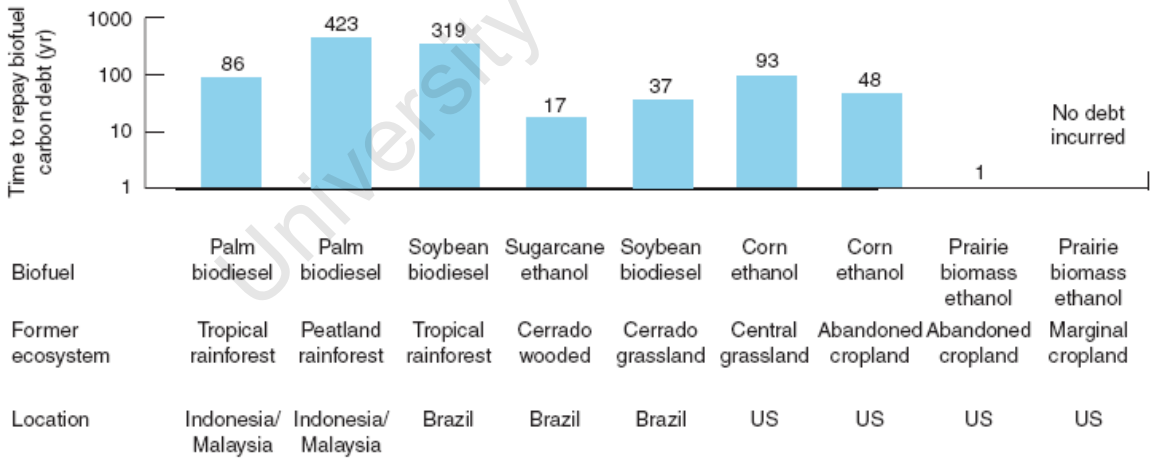


Figure 1-1: Carbon debt repayment times (taken from Fargione *et al.*, 2008)

1.1.2.2. Food price increase

Figure 1-2 shows the increase in bioethanol and biodiesel production over the past 15 years, predominantly since 2002. The rate of increase of bioethanol and biodiesel production from 2002 to 2007 is $8 \cdot 10^9$ litres/yr and $1.5 \cdot 10^9$ litres/yr, respectively. The US and Brazil are the

largest producers of bioethanol, producing $4.3 \cdot 10^9$ litres and $4.2 \cdot 10^9$ litres, respectively in 2005 (EERE, 2008). The EU is the largest producer of biodiesel with Germany and France, producing 625 million litres and 386 million litres respectively in 2002 (IEA, 2004).

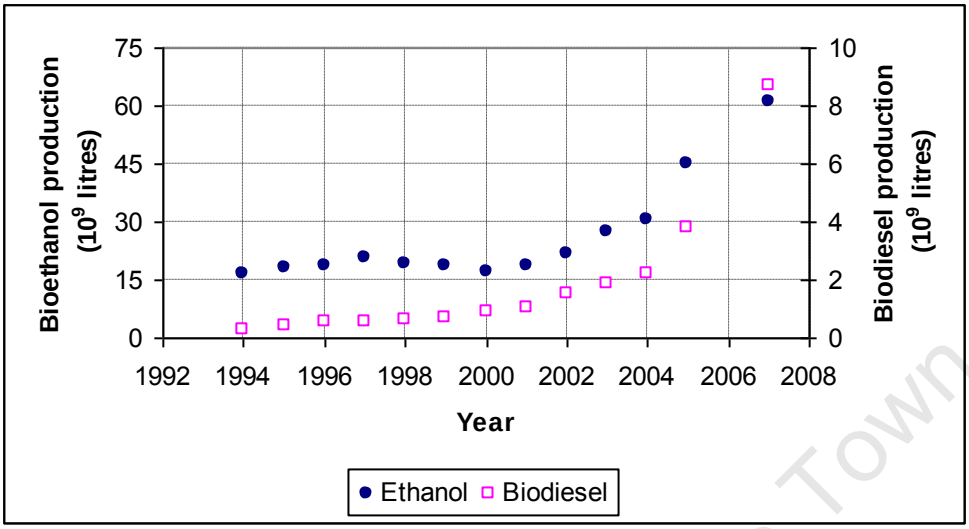


Figure 1-2: Global bioethanol and biodiesel production (adapted from IEA, 2004 and IFPRI, 2008a)

World food prices, especially those of wheat and rice, have risen significantly since the beginning of 2007 as seen in Figure 1-3. Food riots in Mexico due to high prices at the beginning of 2008 were a result of further price increases. There are differing views as to the extent to which the diversion of food and feed crops (such as maize and soya) to energy end-users contributed to the food price increase. In a testimony to the US Senate, the Director-General of the International Food Price Research Institute (IPFRI), stated that 30% of the very recent price hikes in food could be attributed to the increased demand for energy crops (IFPRI, 2008b). A UNEP publication (UNEP, 2008) estimates the contribution to the food price increases at 8% for grains and 17% for seed oils for the 2004 to 2007 period.

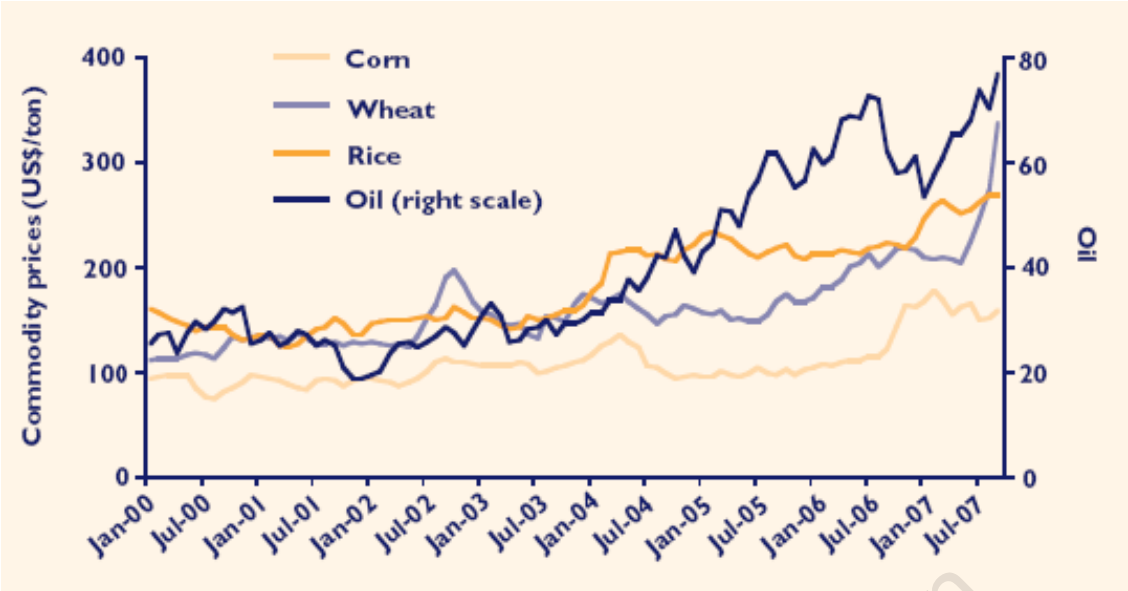


Figure 1-3: Global food prices (IFPRI, 2007)

1.1.2.3. New biofuel legislature in the EU and the USA

In 2008 much discussion in the EU parliament ensued regarding the setting of new biofuel targets. The final target for the transport sector was set at a 10% contribution from biofuels by 2020 of which a 40% share needed to come from hydrogen, electricity and second generation biofuels (European Parliament, 2008). The Commission also proposed that an interim target of 5% contribution by 2010 with 1% from second generation biofuels be set. Energy efficiency was mandated to improve by 20% by 2020 and the carbon emission saving for biofuels must be at least 45% to qualify as biofuel. This would increase to 60% in 2015.

In 2007, the US introduced its Energy Independence and Security Act which includes in the biofuel mix "additional renewable fuel" (essentially second generation renewables). These fuels require a 50% reduction in CO_2 emissions compared to fossil fuels. The target for the supply of the advanced and cellulosic biofuels and biodiesel is 136 Mm^3/yr by 2022 (afdc.energy.gov, 2008). For comparison, current gasoline demand in the States is 385 Mm^3/yr (tonto.eia.doe.gov, 2008).

1.2. Biogenic wastes as second generation renewable fuels

Fuels derived from biogenic waste streams (*i.e.* streams arising from biological sources) are regarded as second generation biofuels as they do not require extraction and processing of a

single plant storage product nor do they compete with food crops. Renewable energy from biogenic wastes is particularly attractive as the waste avoids additional land use and the associated emissions from changing land use.

1.2.1. Types of waste biomass

Biogenic wastes incorporate a variety of stream types. Figure 1-4 delineates the streams considered in this work. The White Paper on renewable energy in South Africa (DME, 2003) identifies residues from agriculture and forestry as well as municipal solid waste incorporating landfill gas capture as potential waste biogenic feedstocks for renewable energy generation. In addition to the lignocellulosic feedstocks, sludges and industrial wastewater streams with high organic loads can also be regarded as potential sources of renewable energy from waste by fermentation to ethanol, anaerobic digestion to biogas or interim accumulation of biomass scavenging the organic residue (Burton *et al.*, 2007).

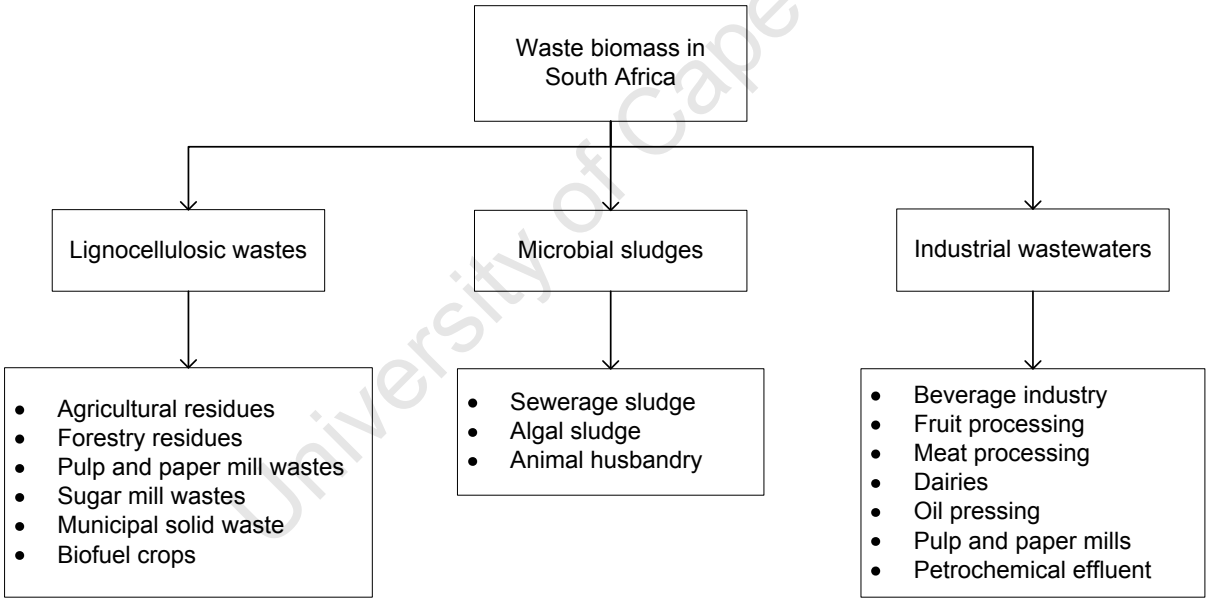


Figure 1-4: Biomass waste sources in South Africa

1.2.2. Characteristics of biogenic waste streams

1.2.2.1. Thermochemical analysis

Waste biomass fuels differ in a number of ways from fossil fuels. Table 1-1 gives the proximate and ultimate analysis for coal and a variety of biogenic waste fuels. Biomass fuels are characterised by

- Lower calorific value than coal (typically 17 to 21 MJ/kg (dry ash-free basis) McKendry, 2002a)
- High volatile content
- High oxygen and hydrogen content (typically up to 30 to 40% oxygen and 5 to 6% hydrogen (Jenkins, 1998))

The ash content of biomass fuels range from less than 5% in wood and 5 to 10% in agricultural crop residues (World Bank technical notes). For biogenic waste fuels such as sewerage sludge or black liquor, the ash content is greater than 30% as seen in Table 1-1.

In comparison, coal generally has a volatile solids content of less than 20% and in the case of anthracite it is negligible (Quaak et al, 1999). The ash in biogenic waste contains a large amount of alkali earth metals, particularly sodium and potassium, which can vaporise at combustion temperatures causing processing issues (Jenkins, 1998).

Table 1-1: Proximate and ultimate analysis of fossil and biomass waste fuels

	Rietspruit coal	Sewerage sludge	Pine woodchips	Corn stover	Bagasse	Black liquor
Proximate analysis (wt% air dried basis)						
Ash	11.7%	35%	2.0%	5.1%	3.80%	40.20%
Volatiles	29.8%	53.5%	57.5%	80.9%	84.20%	46.80%
Fixed carbon	58.5%	11.5%	40.5%	14.0%	12.0%	13.0%
Ultimate analysis (wt% air dried basis)						
C	73.1%	52.3%	47.4%	49.3%	45.90%	29.20%
N	1.8%	7.2%	0.03%	0.70%	1.73%	0.14%
S	0.5%	2.0%	0.1%	0.1%	ND	4.90%
H	0.0%	7.5%	5.2%	6.0%	5.68%	4.40%
O (by difference)	24.6%	31.0%	47.3%	43.9%	46.7%	61.4%
Lower heating value (MJ/kg dry ash-free)	33.6	23.2	17.0	19.0	17.7	17.0
Reference	Paterson, 1997	ECN, 2008	ECN, 2008	ECN, 2008	ECN, 2008	ECN, 2008

1.2.2.2. Biological structure of biomass feedstocks

A biochemical analysis gives the cellulose, hemicellulose and lignin contents of a fuel and is important only in biochemical conversion processes. Lignocellulosic fuels typically contain 35 to 50% cellulose, 20 to 35% hemicellulose and 10 to 25% lignin (Saha, 2003). The primary function of these biological macromolecules is to provide structure to the plant; hence they are tough, recalcitrant materials, resistant to biological degradation. Lignin is not biodegradable at all in current enzymatic hydrolysis systems (McKendry, 2002a). Microbial

sludges resist biological action due to the complex cellular walls of the microbes which make up the sludge (e.g. domestic sewerage sludge or algal sludge) (Speece, 1996).

1.2.2.3. Other

Other factors aside, moisture content, or its inverse, dry solids content, is the single determining factor when choosing an energy conversion process (McKendry, 2002a). Thermal conversion processes typically require moisture content less than 65% to be energetically favourable (Jenkins, 1998) whereas biological conversion processes can utilise high moisture content feedstocks.

The bulk density of waste biogenic fuels affects transport and storage costs and sizing of material handling systems (McKendry, 2002a). Baling and pelleting or briquetting of biomass is often necessary to reduce the transport costs and make handling easier. For example, the bulk density of sawdust is 0.12 ton/m^3 compared to wood pellets which are 0.56 ton/m^3.

1.3. Sustainable energy supply in South Africa

1.3.1. Current energy supply situation in South Africa

Figure 1-5 to Figure 1-8 compare the global primary energy and electricity consumption by feedstock to South Africa. It is clear that South Africa lags behind world trend towards greater inclusion of renewables in the energy mix. The 6% contribution of renewables to total energy consumption in South Africa is primarily from fuelwood and dung for heating in rural areas. However, the practise is unsustainable as it results in deforestation and soil nutrient depletion in the surrounding areas (DME, 2003). The legacy of cheap coal to provide liquid fuel and electricity in South Africa removed the need to diversity our energy mix.

The use of coal as a feedstock to our power plants and their low efficiencies result in high CO_2 emissions per kilowatt-hour of electricity produced. Table 1-2 compares specific CO_2 emissions for various power generating plants

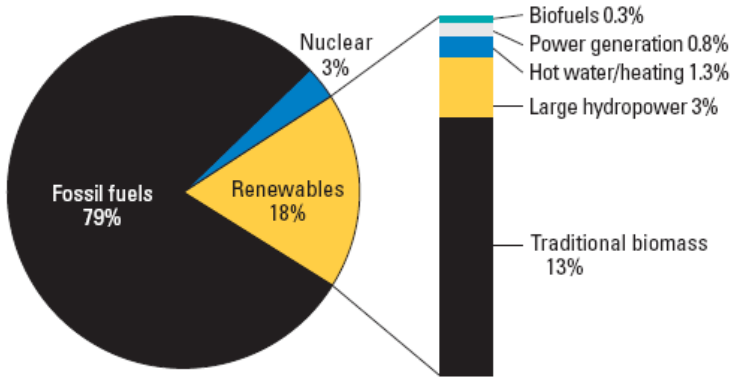


Figure 1-5: Total global energy consumption by feedstock (REN21,2007)

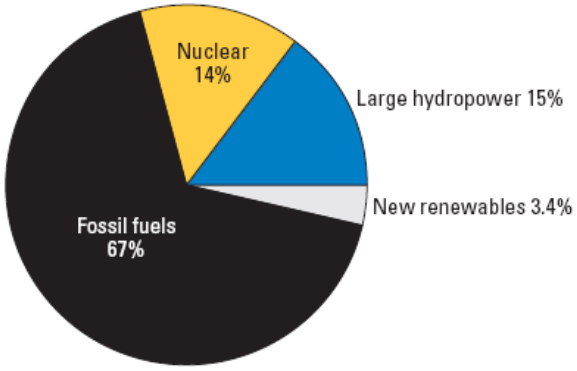


Figure 1-6: Global electricity consumption by feedstock (REN21,2007)

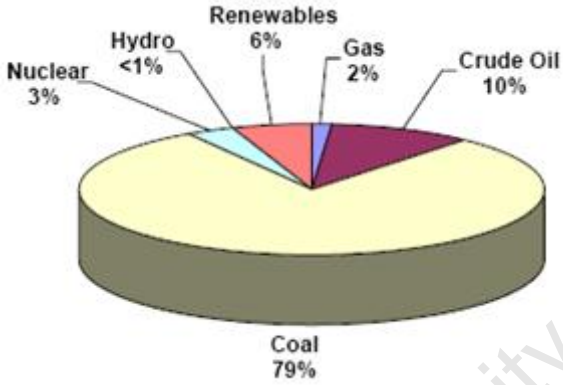


Figure 1-7: Total energy consumption by feedstock in SA (DME, 2005)

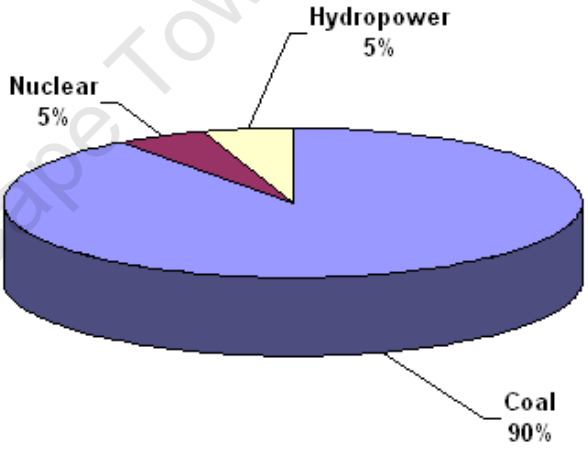


Figure 1-8: Total electricity consumption by feedstock in SA (dme.gov.za, 2008)

Table 1-2: Specific CO_2 emissions from power generating plants

Conversion technology	*Specific CO_2 emissions*	*Reference*
Coal –fired Eskom steam cycle	0.98 kg CO_2/kWhr	Eskom annual report, 2006
Natural gas Combined cycle gas turbine (CCGT)	0.36 kg CO_2/kWhr	Tauschitz and Hochfellner, 2004
Austrian coal fired steam cycle	0.84 kg CO_2/kWhr	Tauschitz and Hochfellner, 2004

1.3.2. Energy efficiency

In terms of South Africa's energy efficiency record, the historically cheap and plentiful power provided by Eskom resulted in inefficient energy practises in the South African industry and our status as an energy intensive nation (SA ranked as the 16th largest consumer of energy in

2001 and our GDP ranked 26th (DME, 2005). The recent energy demand exceeding supply has highlighted the need to revise our energy practises.

Whilst energy efficiency in the EU has been legislated, in South Africa energy efficiency improvements remain voluntary. The Energy Efficiency Strategy released by the Department of Minerals and Energy (DME, 2005) sets a national target for an energy efficiency improvement of 12% by 2015. The improvements are expected to be brought about by economic and legislative means, efficiency labels and performance standards and energy audits. In response, the formation of the National Business Initiative brought 24 companies and 7 industrial associations together to pledge to reduce their energy demand by 15% by 2015 voluntarily. Further business level collaboration has been limited, hence the efficacy of the Strategy without legislation is in question (NBI, 2008).

1.3.3. Increasing the renewable energy share in South Africa

In the White Paper on renewable energy policy for South Africa (2003) a national target of sourcing 10 000 GWhr or 0.8 Million Tone Oil Equivalent (MTOE) from renewable resources by 2013 was set; this equates to 4% of the projected 2013 demand or 4.5% of current demand. Biomass energy, biofuels (as liquid transport fuels), hydropower, wind, solar, geothermal and tidal power are cited as potential renewable energy resources. The Paper specifies that the renewable resources should contribute primarily to electricity supply.

Few estimates of the energy available from biomass energy or biogenic wastes in South Africa exist in literature. The DME provided some estimates in the White Paper (DME, 2004). Williams and Eberhard (1988) and Lynd *et al.* (2003) presented data on the energy potential from lignocellulosic waste streams. Burton *et al.* (2007) reviewed the potential contribution of wastewater sources.

1.4. Problem statement, key questions and objectives

Second generation renewables, including those from biogenic waste streams, may be regarded as a more sustainable energy form than first generation renewables. Further, South Africa needs to increase its share of renewables in the electricity mix. Electricity generation from biogenic waste streams is one possibility which warrants further exploration.

To address the potential for electricity production from biogenic waste streams a series of key questions can be identified:

- What established technologies are suitable for generating electricity from second generation biofuel resources, particularly biogenic waste streams?
- What barriers or technical difficulties exist when processing biogenic wastes and what feedstock pretreatments aid in implementing these technologies to process waste?
- From simple energy balances and operational experience, can the energy yield for different technologies be established, based on intrinsic feedstock characteristics?
- Where more than one technology is available to process a feedstock, can a defined analytical framework assist in technology selection?
- What is the energy potential of biogenic waste streams in South Africa?
- Would realising this potential infringe on alternate uses of the waste?

To address the potential for electricity generation from biogenic waste streams, a series of key objecties can be identified:

1. Identify technologies suitable for processing biogenic waste in South Africa.
2. Develop an energy yield model for technologies suitable for the South African context
3. Validate the model using wastewater treatment plant data
4. Consider the trade-off between the energy cost of feedstock pretreatment and improved energy yield for two commercial microbial sludge pretreatments.
5. Investigate the effect of operational variables on the energy efficiency of combustion of two biogenic waste fuels.
6. Generate a decision-making tool for selecting a technology based on feedstock characteristics.
7. Determine the potential energy from biogenic waste in SA using the above tools and compare these estimates to previous studies.
8. Consider the alternate uses of biogenic waste streams or, if applicable, their current use and the energy efficiency thereof to enable a realistic estimation of biogenic electricity potential in South Africa.

1.5. Thesis structure

Chapter 2 considers technologies which may be used to generate electricity from biogenic waste streams. Specifically the process, reactor types and current installations, challenges and feedstock preparation techniques for combustion (Section 2.2), gasification (Section 2.3) and anaerobic digestion (Section 2.4) are considered.

In Chapter 3, case studies of three wastewater treatment plants which feature different technologies for energy recovery are presented. While these data were sourced from the UK, they are not regionally dependent and so have relevance to the South African case. The selection of the UK data resulted as South African data of equivalent detail was not available. Plant operation observations are presented to highlight the challenges of these systems and their potential. The ability of the technologies to deliver heat and power are compared. An energy yield model is derived and is compared to plant data to investigate its ability to assist decision making.

Chapter 4 considers the effect of characteristics specific to biogenic waste on the processing of these fuels. Two studies are presented: the effect of recalcitrant material on biogas yield in anaerobic digestion (Section 4.1) and the effect of high volatile content on combustion in fluidised beds (Section 4.2).

In Chapter 5 a technology selection methodology is presented to assist in estimating the potential energy yield from biogenic waste streams in South Africa. Alternate uses and current practice are compared with best practice, allowing realistic energy estimates to be prepared. The conclusions and significance of the study are presented in Chapter 6.

Chapter 2 Technologies suitable for heat and electricity generation from biogenic waste streams

2.1. Introduction

Biogenic waste streams may be treated thermochemically through combustion or gasification or biologically through anaerobic digestion to raise steam or produce a combustible gas. In either case, these can in turn be used to generate electricity. Biomass-based systems are, conceivably, the only renewable energy source apart from hydro-power which could generate base load electricity. Figure 2-1 shows the biomass conversion routes investigated in this chapter. Biogenic waste streams may also be processed to produce liquid transport fuels such as cellulosic ethanol or biodiesel; however, this is beyond the scope of this work.

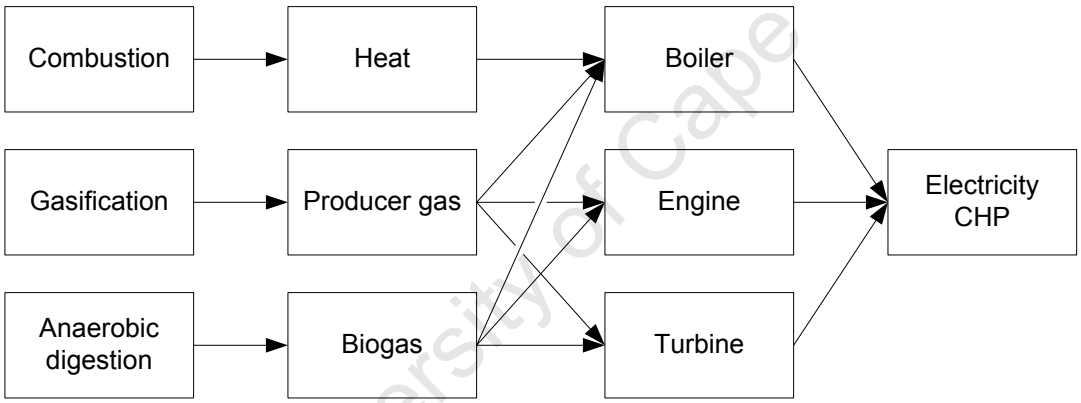


Figure 2-1: Energy conversion routes for biomass

2.2. Combustion

2.2.1. Process

Electricity generation based on biomass steam cycle power plants is a commercial reality, particularly in the USA and the Scandinavian countries (Rosillo-Calle *et al.*, 2000). The technology thus presents the least risk to investors (Bridgewater, 2003). Combustion, unlike the biochemical energy conversion processes, is essentially non-selective in its use of the feedstock, and reduces the whole fuel to simple products (Jenkins *et al.*, 1998). Biomass can be combusted on its own or co-fired with coal to raise steam. Power cycle exhaust steam may be used plant-wide for process heat. Figure 2-2 shows the location of the boiler tubes within

the furnace. The relative simplicity of the process is negated by low thermal efficiencies. Efficiencies of 15 to 20% (Bridgewater, 2003; McKendry, 2002b; Rosillo-Calle *et al.*, 2000) are found in small biomass boilers. This may increase to 30% for larger boilers (Bridgewater, 2003); or to 40% for co-firing in coal power plants (Mc Kendry, 2002b). Low thermal efficiencies can be attributed to different optimal capacities for biomass boilers and steam turbines. Biomass boilers are limited in size to prevent prohibitive feedstock haulage costs whereas steam cycles are ideally large to realise economies of scale. As economies of scale are not realised for the steam cycle less technology is incorporated into the plant to offset costs, resulting in lower efficiencies (Rosillo-Calle *et al.*, 2000). Steam cycle efficiency can be improved by increasing the boiler pressure and using condensing steam turbines. An example from the sugar industry is shown in Figure 2-3.

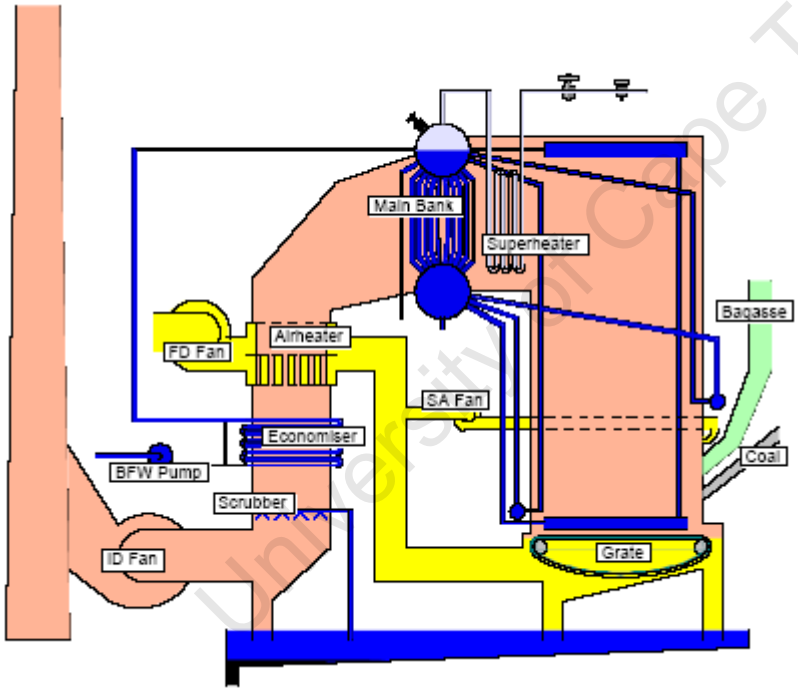


Figure 2-2: Boiler configuration (taken from Wienese, 2001)

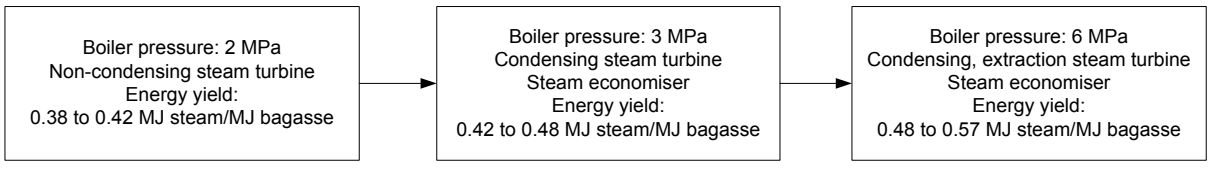


Figure 2-3: Measures to improve thermal efficiency (adapted from Beeharry, 1996)

The amount of heat released from a biomass fuel is given by the Lower Heating Value (LHV) or the Higher Heating Value (HHV). LHV measures the enthalpy change with the water vapour exiting the system uncondensed. HHV measures the enthalpy change with the water vapour exiting as a liquid, and is the value measured by bomb calorimetry. LHV applies to conventional boiler systems; HHV applies when the heat of condensation is recovered, *i.e.* when a condensing steam turbine is used.

Combustion of biomass proceeds via drying and devolatilisation, followed by rapid gas-phase oxidation of the volatiles and char combustion (Nussbaumer, 2003; Jenkins *et al.*, 1998) as shown in Figure 2-4. The higher H and O contents in biomass fuels result in lower heating values than fossil fuels, as demonstrated in Figure 2-5. However, the high oxygen content also reduces the stoichiometric air requirement and thus amount of inert N_2 to be heated to reach adiabatic flame temperature (Jenkins *et al.*, 1998).

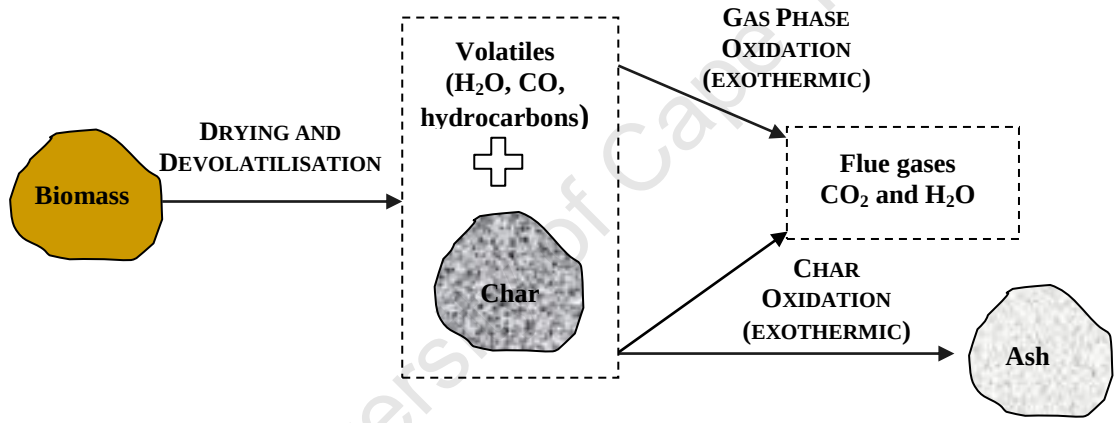


Figure 2-4: Process of combustion of biogenic feedstocks

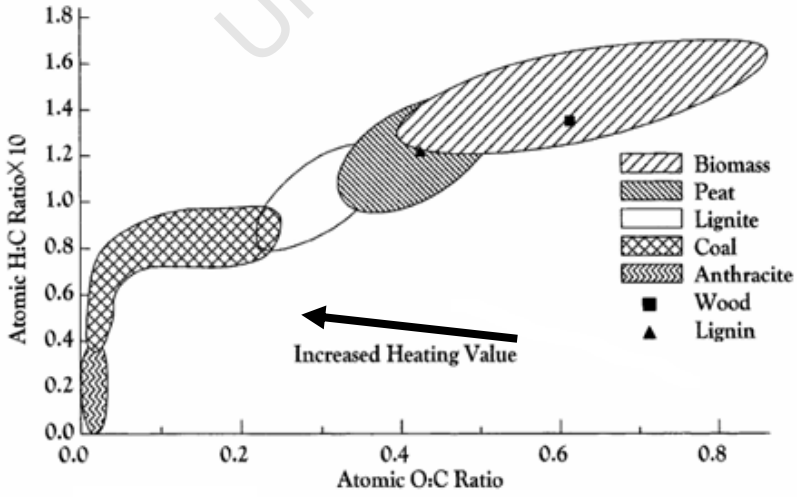


Figure 2-5: Van Krevelan diagram for solid fuels (taken from McKendry, 2002b)

2.2.2. Reactor types and current installations

Combustors used for biogenic waste combustion include fixed and fluidised bed designs. Fixed beds are the oldest design and offer simpler operation (Quaark *et al*., 1999). Fixed bed combustors are further described in Table 2-1. Fluidised beds are regarded as the best reactor type for processing waste materials (Nussbaumer, 2003; Anthony, 1995; Werther and Ogada, 1999; Koorneef, 2007; Bain *et al.*, 1998). Reasons cited include:

- Intimate mixing results in high heat transfer in the bed and results in uniform bed temperatures
- The thermal mass of the bed material maintains temperature and promotes complete carbon conversion in the bed
    - This allows for operation at relatively low temperatures and with low excess air
    - This also allows for low grade fuels to be burnt. an auxiliary fuel
- Low operating temperatures mean that little thermal NO_x is produced
- Opportunity to reduce SO_x emissions with the use of sorbents as bed material

A fluidised bed combustor operates by passing a stream of air through a bed of granules; typically sand. The air enters the bed from the plenum chamber and passes through the distributor plate. The minimum fluidisation velocity (U_{mf}) is the velocity at which the bed changes from fixed to fluidised. The superficial air velocity determines the type of fluidisation, which progresses from bubbling bed fluidisation to slugging bed fluidisation to particle entrainment with increasing air velocity as shown in Figure 2-6.

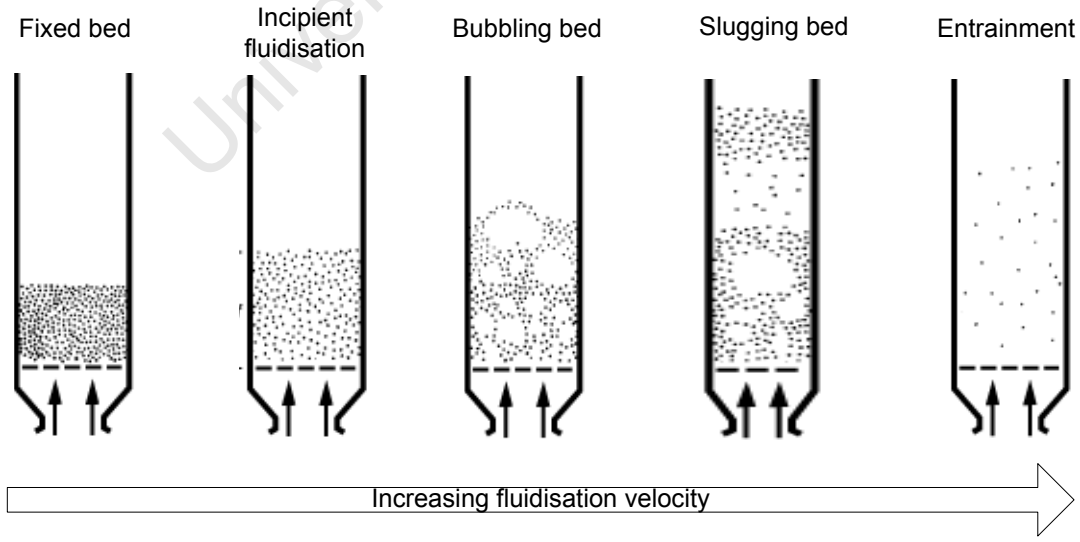
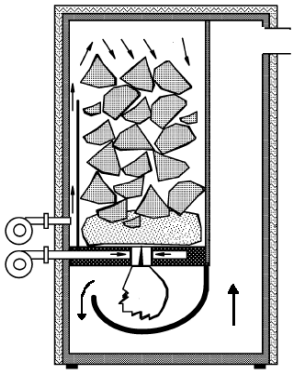
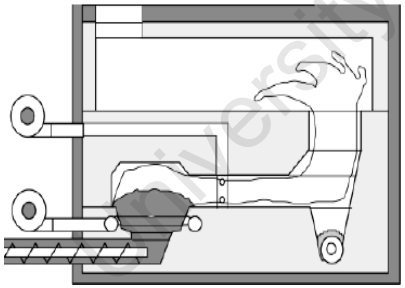
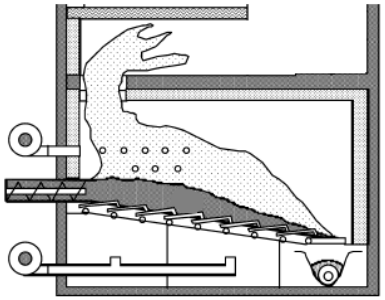
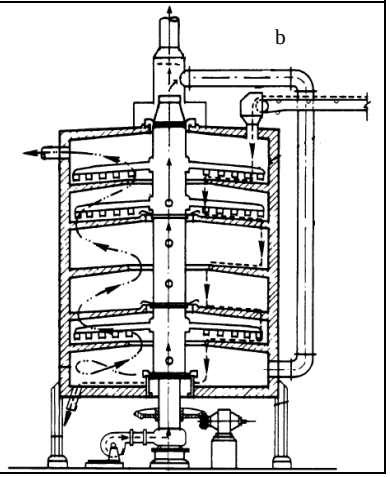


Figure 2-6: Flow regimes for fluidised beds (Davidson *et al.*, 1977)

Table 2-1: Combustion furnace reactor types (adapted from Nussbaumer, 2003)

	Downdraft boilers	Understoker boilers	Moving grate and spreader-stocker boilers	Multiple hearth furnace[a]
Process description	Primary air moves downward through successive zones of drying, devolatilising and combustion. Secondary air is added to the post-combustion chamber to burn remaining volatiles and tars	Fuel enters the combustion chamber via an auger. Primary air enters underneath the combustion chamber and secondary air enters in the post-combustion chamber.	Primary air enters below the grate. Fuel is fed onto the grate via an auger or, in the case of a spreader-stoker, via a high speed rotor which flings the fuel evenly onto the grate. Secondary air is introduced above the grate. The fuel on the grate moves through sucessive zones of drying, devolatilisation and combustion	Fuel is fed to the top hearths (shown by the arrows on the right) and air is introduced at the bottom (shown by the arrows on the left). Fuel is passed to lower hearths by the rotating arms through successive zones of drying, devolatilisation and combustion.
Application	Achieves complete burnout and high efficiencies in small-scale units when primary and secondary air is employed	Used mostly for wood chips and similar fuel with low ash and moisture content.	Can be applied to fuels with higher ash and moisture content.	The design features good internal energy usage as the incoming feed is directly heated by the hot flue gases.
Reactor diagram showing air flow				 b

a. Werther and Ogada, 1999
b. EPA, 2008

In a bubbling fluidised bed, gas bubbles rise through the bed at a velocity U-U_{mf}. In the interstitial spaces between the granules, gas moves upward at velocity U_{mf}. These two phases are known as the bubble phases and particulate phase respectively (Davidson *et al.*, 1977). The combustion of gases (or volatiles) in a fluidised bed takes place rapidly in the bubble phase, as the reaction proceeds via a radical mechanism which is suppressed in the particulate phase by the presence of solid surfaces (Hayhurst and Parmar, 1998). Work by Turnbull and Davidson (1984), Hesketh and Davidson (1991) and Ogada and Werther (1996) shows a bed temperature dependence on the location of volatiles combustion, with higher temperature promoting in-bed bubble phase combustion. This is further explored in section 4.2.

Circulating fluidised beds operate at air velocities great enough to entrain the bed material particles. The entrained bed is separated from the flue gases in the disengagement drum and returned to the combustor via a downcomer. Fuel is introduced to the bottom of the bed. Depending on the fuel particle size, particles will immediately be entrained and burn during circulation or burn in suspension for before becoming entrained. The circulating fluidised bed has more fuel flexibility than the bubbling fluidised bed as the rate of material circulation can be used as an additional control (Quaark *et al*., 1999).

Both bubbling bed and circulating fluidised bed technologies may be regarded as mature technologies for co-generation and industrial sized applications. Bubbling fluidised beds are typically used for industrial applications less than 25 MW_e (Koornneef *et al.*, 2007). Owing to their ease of scale up, circulating fluidised beds find application in the power generating sector, where units greater than 75 MW_e are typical. Data from Koorneef *et al*. (2007) shows that 70% of installed fluidised bed are circulating beds; this represents 86% of installed capacity. Economies of scale play a significant role with scale factors for biomass fluidised bed combustion projects ranging from 0.6 to 0.8 (Koornneef *et al*., 2007).

Fluidised bed combustion of waste streams has not found widespread industrial application in South Africa to date. Recently eThekwini Municipality installed a fluidised bed drier followed by a fluidised bed boiler to dry and incinerate sludge pellets. However, the unit did not provide stable operation and is currently not running (Dildar, 2008). A local company, Carbon and Environmental Options, in partnership with the Eden municipality ran trials in fluidised bed combustors using waste wood chips from the logging operations in the area. The

use of fluidised beds in the project was rejected on a cost basis and the boiler now under consideration is a spreader-stoker type (Eleftheriades, 2008). A coal and biomass co-fired fluidised bed combustor has been successfully installed at a food processing factory in Escourt, KZN, to dispose of coffee grounds (North *et al.*, 2005).

2.2.3. Feedstock preparation

2.2.3.1. Lignocellulosic feedstock preparation

Preparation of lignocellulosic waste for combustion or gasification is well established. The process is briefly outlined in Figure 2-7. Communition with associated classification reduces material to particle sizes of 20 to 80mm (McKendry, 2002c). To allow combustion of lignocellulosic materials to proceed endogenously, drying is necessary to reduce moisture contents to less than 50% to 60% (*i.e.* dry solids content of 40% to 50%) (Nussbaumer, 2003; McKendry 2002b; Jenkins, 1998). Drying to 10 to 15% for gasification systems is considered acceptable in terms of costs and energy output for gasification as it prevents condensation problems and reduces tar content (Bridgewater, 1995). Leaching in water prior to gasification may be used to reduce alkali metal content of the feed and mitigate operational problems such as bed agglomeration (McKendry, 2002c).

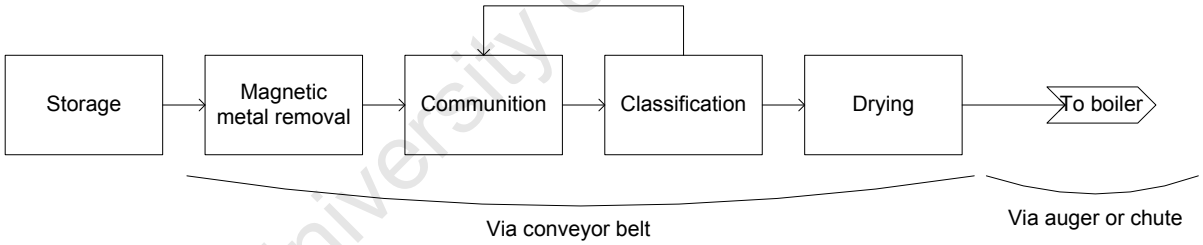


Figure 2-7: Feed preparation for lignocellulosic materials (adapted from Bridgewater, 1995)

2.2.3.2. Sludge feedstock preparation

Sludges, and in particular sewerage sludge, exist as a suspension containing 0.25 to 12% dry solids (Werther and Ogada, 1999). The moisture associated with the dry solids and the means of its removal is described in Table 2-2. The material balance and associated units used to achieve this are shown in

Table 2-3. Mechanical dewatering techniques commonly found on wastewater treatment plants include centrifuges, belt filter press and filter plate presses. With the use of flocculants, these units produce cake with a solids content of 20 to 40% (Werther and Ogada, 1999). Dewaterability, especially of biological sludges, is further improved with *inter alia* thermal

conditioning as this ruptures the cell water and releases the intracellular water. Comparative capital expense and operating expenses of the dewatering units are given in Table 2-4.

Table 2-2: Types of water found in sludge (adapted from Werther and Ogada, 1999 and Vesilind, 1994)

Moisture association	***Description***	***Removal technique***
Free water	Water not associated with or influenced by solid particles	Thickening
Interstitial or floc water	Water trapped in the crevices and interstices of flocs and organisms	Mechanical dewatering
Capillary or vicinal moisture	Layers of water held to the particle surface by hydrogen bonding	Mechanical dewatering with chemical conditioning
Intracellular water	Water inside the organism	Water only released after cell disruption

Table 2-3: Materal balance and unit operations for increasing dry solids content of sludge

Process	Thickening		Dewatering	
Unit operation	Gravity thickener	Gravity belt thickener	Belt filter press; filter plate press; solid bowl centrifuge	
Dry solids content of sludge	1%	6%	15%	28%
Total mass (kg/kg dry solids)	100	17	7	4

Table 2-4: Operating and capital expenses of dewatering equipment

	Capital expense [a]	***Operating expense*** [a]	***Energy consumption***
Belt press	Low	Low	30 to 60 MJ/ton dry solids
Filter plate press	High	High	Similar to above
Solid bowl centrifuge	Medium	Medium	100 to 300 MJ/ton dry solids

a.) Bane, 2000
b.) adapted from Wang *et al*., 2007

(1) Gravity belt thickeners

Horizontal gravity belt thickeners consist of a continuously rolling belt onto which sludge is evenly introduced at one end. Thickening occurs by gravity as the water falls through the porous belt. The thickened sludge is discharged at the other end of the belt. The belts are self-cleaning as the water dropping through the belt at the top fall onto the belt below. The unit is good for coarse fast settling solids (McCabe *et al.*, 2005). With the addition of polymer flocculants, gravity belt thickeners increase the dry solids content from 1% to 6% dry solids content (Fountain, 2008). Although this unit is strictly speaking a thickener and not a dewatering unit, it is included here for completeness as it is used upstream of the dewatering units.

(2) Belt filter press

Belt filter presses comprise of a continuously rolling belt onto which the sludge is fed. Following a section of gravity thickening, the sludge is then squeezed between two opposing porous cloth belts. The belts are rolled between successively smaller rollers which increase the pressure and force the water through the belt. The sludge is scraped off the belt into a collection bin by scraper blades. Belt filter presses increase the solids content from about 5% to 15% to 28% depending on the proportion of biological sludge in the feed. The unit is the preferred dewatering technique in the US. Its main disadvantage is sensitivity to wide variations in sludge characteristics (Tchobanoglous *et al.*, 2003).

(3) Filter plate press

A filter plate press is operated in batch mode. It consists of a number of plates, recessed on both sides, supported on a frame. A filter cloth covers each plate. Sludge is introduced to the compartments between each plate under 3 to 10 bar pressure. Water is forced through the filter cloth into a discharge pipe. Once sufficient water has been removed the pressure is released and the plates are separated. The wet cake is dropped into a collection bin (McCabe *et al.*, 2005). Discharge solids contents of 48 to 70% cited (Tchobanoglous *et al.*, 2003); however operational difficulties, such as blinded filter cloth, can result in discharge solids contents of 20 to 25% (Fountain, 2008). Disadvantages of the unit include excessive downtime for equipment maintenance and high labour costs (Tchobanoglous *et al.*, 2003)

(4) Solid bowl centrifuge

A solid bowl centrifuge consists of a horizontally mounted centrifuge bowl inside which a helical scroll, rotating slightly slower than the bowl, moves the solids to one end of the bowl. "Centrate" or liquor is withdrawn from the other end (McCabe *et al.*, 2005). Discharge solids content ranges from 10 to 30% depending on the amount of biological sludge (Tchobanoglous *et al.*, 2003). Disadvantages of the unit include high power costs (typically three times more than belt filter presses (Fountain, 2008)).

2.2.4. Summary

Combustion of biogenic waste fuels, particularly woody biomass and sewerage sludge is well-established technology. The units involved in the preparation of fuel prior to combustion are

also well established. It has been highlighted that fluidised bed combustors offer a number of advantages over fixed bed designs. However, fixed bed designs remain the simpler option. In South Africa industrial experience in fluidised bed technology exists, notably the co-fired unit at Escourt. The main drawback of the use of biomass-only combustion systems are the low thermal efficiencies achieved, especially is small scale units.

2.3. Gasification

2.3.1. Process

Gasification is a thermal process which yields a combustible gas, called "fuel gas" or "producer gas" with a heating value of 5 to 16 MJ/Nm^3, depending on gasifying agent used. Gasification occurs in an atmosphere of sub-stoichiometric oxygen requirement. Fuel particles are heated and release their volatile components. The carbon in the particle is consumed via combustion and gasification reactions leaving behind the ash, as shown in Figure 2-8. These reactions may proceed sequentially or simultaneously depending on heating rate.

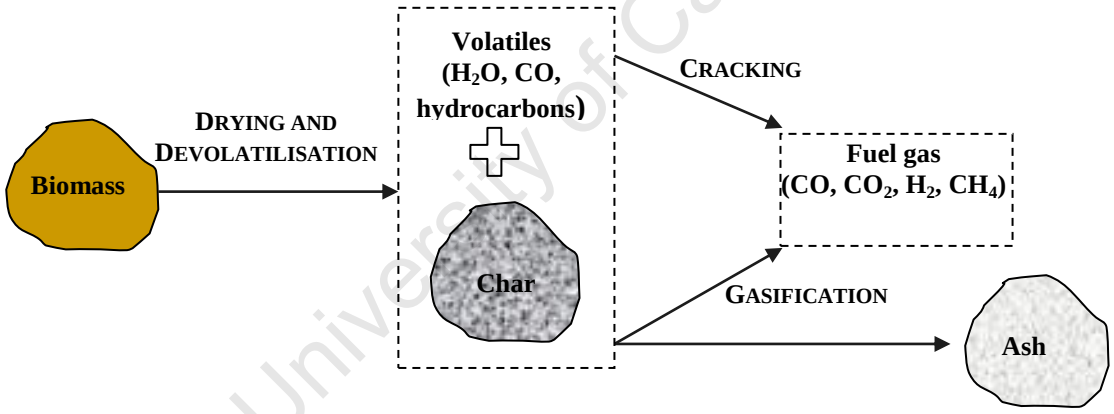


Figure 2-8 - The process of gasification

Important gasification reactions are shown in Table 2-5. It can be seen that those converting solid carbon to valuable fuel gas species are endothermic, equilibrium reactions (Scott, 2004). These reactions are also the rate limiting reactions (Bridgewater, 1995). To run a gasifier adiabatically, the endothermic and exothermic reactions have to be coupled so that the maximum amount of CO is produced while enough carbon is combusted to maintain temperature. This is controlled by the equivalence ratio, which is the amount of O_2 fed to the amount of O_2 required for complete combustion. Scott (2004) found the optimal equivalence air ratio for sewerage sludge gasification to be 0.35. The range is 0.2 to 0.4, with higher

values resulting in decreased tar formation but lower calorific gas due to dilution with nitrogen (Narvaez, 1996). The hot-gas efficiency of gasification is 95 to 97%, meaning that most of the energy in the feedstock is converted into product (Bridgewater, 2003).

Table 2-5: Important gasification reactions (taken from Scott, 2004)

Heterogeneous reactions	ΔH^{o}_{298} (kJ/mol)	
$C + ½ O_2 \rightarrow CO$	-111	Combustion
$C + O_2 \rightarrow CO_2$	-394	Combustion
$C + H_2O \leftrightarrow CO + H_2$	+131	Gasification by H_2O
$C + CO_2 \leftrightarrow 2CO$	+173	Gasification by CO_2
$C + 2H_2 \leftrightarrow CH_4$	-74.9	Hydrogasification
Homogeneous reactions		
$CO + H_2O \leftrightarrow CO_2 + H_2$	-41.2	Water-gas shift reaction
$CO + 3H_2 \leftrightarrow CH_4 + H_2O$	-206	Methanation
$CO + ½ O_2 \leftrightarrow CO_2$	-283	Combustion
$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$	-803	Combustion

Gasification affords higher efficiencies than steam cycle combustion power plants through Biomass Integrated Gasification Combined Cycle Gas Turbines or BIG-CCGT systems (see Figure 2-10). Efficiencies of BIG-CCGT systems can reach 45 to 50% in 100 MW_e plants (Rosillo-Calle *et al.*, 2000; Demirbas, 2004). Smaller BIG-IGCC plants (less than 10 MW_e) achieve efficiencies of 30% (Bridgewater, 1995). Alternatively, the producer gas could be used on a small scale (150 kW_e to 5 MW_e) in internal combustion engines to generate heat and power. However, these systems are sensitive to tar and as gas cleaning technology is not well-developed yet, costs may be prohibitively high (Rosillo-Calle *et al.*, 2000). Efficiencies are also low at 30% (Bridgewater, 1995). Pressurised gasifiers negate the need for fuel gas compressors and can provide efficiencies of 45% in 30 to 50 MW_e plants but capital and running costs are higher than atmospheric units (Klimantos et al, 2008). The thermal efficiencies of gasification and combustion systems as a function of capacity are shown in Figure 2-9.

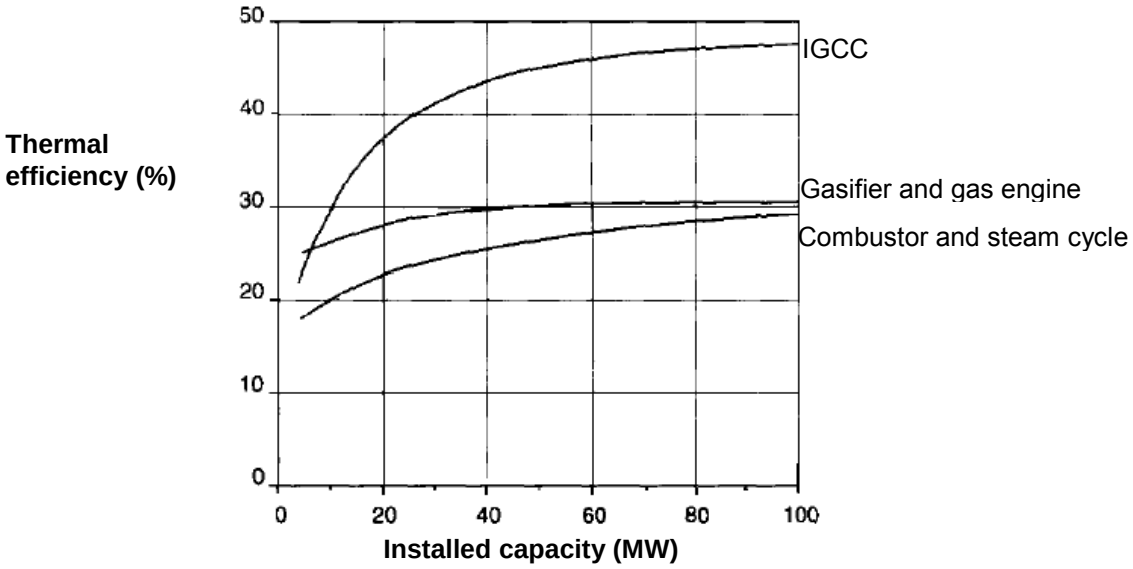


Figure 2-9: Efficiencies of thermal processes as a function of installed capacity (Bridgewater, 2003)

Air-blown gasifiers are typically used; however, the low calorific value of the gas prevents its use in unmodified gas turbines or internal combustion engines. Modifications include adaptation of the fuel nozzles in internal combustion engines and fuel manifold in gas turbines (Rosillo-Calle *et al.*, 2000). Alternatively, "de-rating" lowers the operating temperature of the gas turbine or internal combustion engine which leads to lower efficiencies (Rodrigues, 2003). The use of oxygen blown gasifiers results in a higher calorific gas (10 to 12 MJ/Nm^3), but is negated by the cost of providing the oxygen (Bridgewater, 2003).

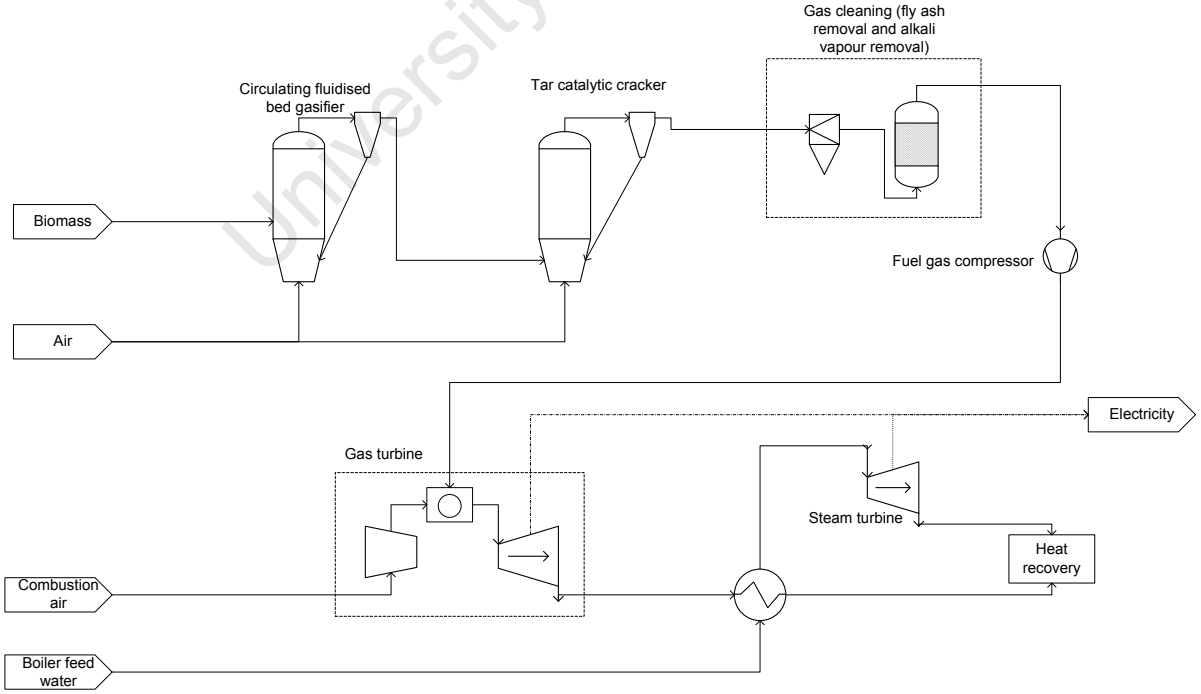


Figure 2-10: BIG-CCGT system (adapted from Rosillo-Calle *et al.*, 2000)

2.3.2. Reactor types and application

The main designs of gasifier are: updraft fixed bed, downdraft fixed bed, bubbling fluidised bed and circulating fluidised bed; the main features of these are highlighted in Table 2-6. A review of major commercially gasifier manufacturers in the EU, USA and Canada showed that 75% of offered designs were downdraft and 20% were circulating fluidised beds (Bridgewater, 2003). Despite the advantages offered by gasification, little information about costs, emissions, efficiencies and turn down ratios exists for gasifier designs. This highlights the shortage of large scale operational experience of biomass gasification.

The feedstocks used for small scale gasifiers, according to a survey presented in Rosillo-Calle (2000), is wood (56%), agricultural residues (27%) and other (16%). A similar survey of larger BIG-CCGT plants (most of them are not yet commissioned) showed that 5 of the 8 used some form of wood and the remainder used agricultural residues (Rosillo-Calle, 2000).

Experience in biomass gasification in South Africa is limited. Sasol carried out co-firing tests on their fixed-bed, dry-bottom Lurgi gasifier at Secunda (Tait, 2006). A mixture of bark and wood fibre pulp was added to the usual coal mix to make up 9% of the feed. For stable operation, the steam:oxygen ratio had to be increased which resulted in a higher than usual production of CO_2 (*i.e.* lower carbon efficiency). At conditions optimal for coal gasification, unstable operation occurred; with higher flue gas temperatures experienced. Biomass-only systems are not being actively pursued at Sasol for the following reasons (Tait, 2006)

- Biomass gasification and gas cleaning technology needs further development
- High cost of delivered biomass feedstock due to low density of biomass
- Low fixed-carbon content of biomass compared to coal. This limits desirable heterogenous reactions

Table 2-6: Gasifier reactor types

Reactor type	*Gas quality* [a]	*Application* [c]	*Example* [c]	*Output*
Updraft gasifier	High tar content	Small scale fuel gas production for use in boiler	A few Scandinavian district heating systems	4-6 MW_{th}
Downdraft gasifier	Lowest tar content [b], particulate content high	Best for small scale electricity production	Numerous small scale applications in USA, Scandinavia, India and China	< 1 MW_e
Bubbling fluidised bed	Low tar content	Small scale CHP	Not many installed units	< 5 MW_e
Circulating fluidised bed	Low tar content	Best for large scale electricity production	Varnamo pressurised gasifier demonstration plant, Finland (decommissioned in 2000) [d] ARBRE demonstration plant (project never reached operation) [e] TPS/Electrobras [f] (project abandoned in 2004 [g]) SilvaGas (Vermont, USA) (run from 1998 to 2001[h]) Greve-in-Chianti TPS gasifier (commissioned in 1993, renovation began in 1997 and completed in 2000)[j]	6 MW_e 8 MW_e 32 MW_e 15 MW_e [i] 6.7 MW_e [j]

a.) Bridgewater, 2003
b.) McKendry, 2000c
c.) Rosillo-Calle, 2000
d.) Klimantos et al, 2008
e.) Piterou et al, 2008
f.) Waldheim and Carpentieri, 2001
g.) tps.se, 2007
h.) http://rentechinc.com/silvaGas.php, 2009
i.) calculated from data in Paisley and Overend, 2002
j.) Granatstein, 2003

2.3.3. Challenges

2.3.3.1. Tar conversion

Gas turbines and internal combustion engines are sensitive to the presence of small amounts of tar. As such, tar conversion remains the most significant challenge facing electricity from biomass gasification projects (Bridgewater, 2003; Demirbas, 2004) Moreover, tar formation lowers the hot gas efficiency of the gasifier. Tars are formed from the semi-volatile and non-volatile compounds released by the biomass fuel (Brage, 1996). Tar formation and tar conversion methods are generally independent of fuel type (Milne and Evans, 1998 and Fryda *et al.*, 2007).

(1) Types of tar

Definition of a standard set of tar types is required when studying tar conversion (Kiel *et al.*, 2004). Experts at the EU/IEA/US-DOE meeting on tar measurement protocol, (Brussels

1998) agreed to define tar as all organic contaminants with a molecular weight higher than benzene (Nordgreen, 2004). A more detailed classification, adopted at the World Biomass conference in Italy in 2004, is provided in Table 2-7. In their comprehensive report on the nature, formation and conversion of tars, Milne and Evans (1998) characterise a comprehensive list of compounds into primary, secondary and tertiary tars, as detailed in Table 2-7. Primary tars can be thermally cracked at 800°C; secondary tars can be cracked at 1000°C. Thermal cracking has the disadvantage that at high temperatures tertiary, refractory tar species form. Milne and Evans (1998) cite the tolerance limits for condensable tars in gas turbine applications as 0.05 to 0.5 ppm. The limit for non-condensable tars is higher.

Table 2-7: Five classes of tar (adapted from Kiel *et al.*, 2004 and van Paasen and Kiel, 2004)

Name	Tar compounds	Identification	Impact	Milne and Evans class
Class 1	Fraction heavier than coronene (GC-undetectable)	Total gravimetric fraction less GC-detectable fraction	These tars condense at high temperature and low concentration	Tertiary
Class 2	heterocyclic aromatics	pyridine, phenol, cresol, quinoline	Exhibit high water solubility, easily cracked.	Primary
Class 3	aromatics (1 ring)	benzene, xylene, styrene, toluene	These are light hydrocarbons not important in condensation or water solubility issues	Secondary
Class 4	light PAH compounds (2-3 ring)	naphthalene, biphenyl, flourine	These compounds condense at relatively high concentrations and intermediate temperatures	Tertiary
Class 5	heavy PAH compounds (4-7 ring)	fluoranthene, pyrene up to coronene	These compounds condense at relatively high temperature and low concentration.	Tertiary

(2) Methods of tar conversion

Two broad methods for tar conversion exist: in-bed technology and downstream reforming. Catalytically active beds, containing for example dolomite and olivine sand, can be used to reduce tar formation (Rapagna *et al.*, 2000). However, materials such as dolomite are subject to attrition and the formation of fines may clog up downstream filters. Olivine sand has shown greater mechanical strength than dolomite but a lower conversion efficiency (94% reduction in flue gas tar content compared to 99% reduction achieved with dolomite (Rapangna et al, 2000). Downstream reforming processes generally use a fluidised bed of metallic, tar cracking catalyst. Nickel steam reforming catalysts have shown good activity but are subject to de-activation in the presence of oxygen (Baker *et al.*, 1987). Data from Baker et al (1987) indicates a reduction in tar content of 54% when using a nickel catalyst. Metallic iron was also shown to be catalytically active (Nordgreen *et al.*, 2006), achieving a reduction in tar content of 89% in a downstream fluidised bed operated at 800°C.

2.3.3.2. Bed agglomeration

Bed agglomeration, due to ash softening, is a problem in fluidised bed combustors and fluidised bed gasifiers. In the latter, it is aggravated as the reducing atmosphere lowers the ash agglomeration temperature. Resolving the problem requires prior knowledge of the ash melting temperature, allowing prediction of bed agglomeration before bed defluidisation and process modifications to raise the ash melting temperature.

(1) Bed agglomeration process

The major elements present in ash are: Na, K, Ca, Mg, Al, Fe, S, Cl and P. These may exist as metal oxides (e.g. Na_2O and K_2O) or bound to the organic matter of the biomass (e.g. S, Cl and Ca). Inorganic material bound to the organic component of the biomass volatilises at lower temperatures than those existing as oxides (Zevenhoven-Onderwater *et al.*, 2001). Bed agglomeration occurs when gaseous phase alkali metals vapourise and then condense onto bed material particles. The resulting sticky residue can "glue" adjacent particle together, ultimately resulting in bed defluidisation. The process of bed agglomeration is presented in Figure 2-11.

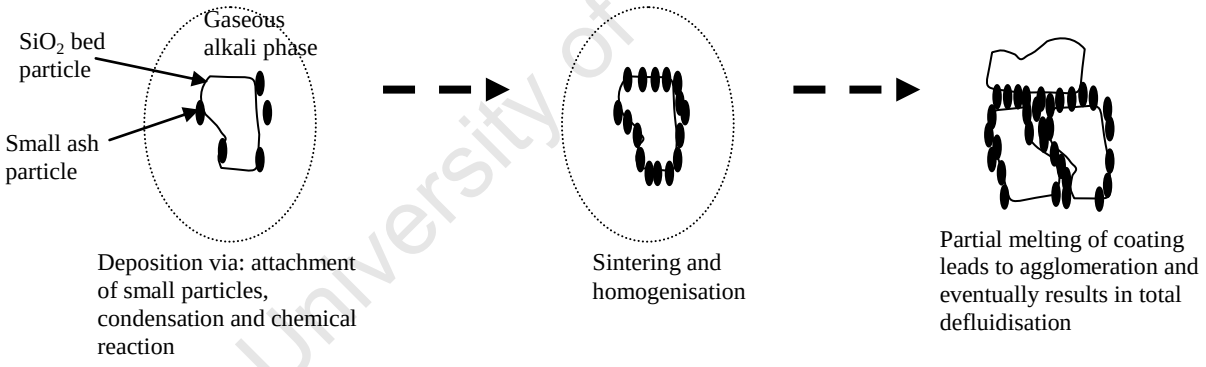


Figure 2-11: Process of bed agglomeration (Ohman and Nordin, 2000)

(2) Determining ash melting temperature

The ASTM ash fusion test is generally considered to be a poor indicator of ash fusion temperature, but it is still widely used as no simple, reliable alternative has as yet been developed (Natarajan *et al.*, 1998; Zevenhoven-Onderwater, 2001). The ASTM ash fusion test cannot factor in the influence of bed material or the influence of alkali vaporisation. It is not run under conditions found in the gasifier and is observational in nature, *i.e.* the melting temperature is determined visually.

Natarajan *et al.* (1998) compared the ash fusion temperature predicted by the ASTM ash fusion test to those obtained in a bench scale fluidised bed combustor. They found the ASTM test to over-predict the ash fusion temperature in all feedstocks investigated. Similarly, Skrifvars *et al.* (1999) compared the bed melting temperatures predicted by the ASTM ash fusion test, a compression strength sintering test and a bench scale fluidised bed combustor. The results showed that in all cases the ASTM test predicted fusion temperatures 200 to 400°C higher than the bench scale combustor. In most cases the compression strength sintering test predicted ash fusion temperatures 50°C lower than the bench scale combustor.

Equilibrium models have been used to determine the ash melting temperature. Zevenhoven-Onderwater *et al.* (2001) carried out detailed thermodynamic equilibrium calculations to calculate the amount of condensed phases present under reducing conditions and at different temperatures. The advantage of their model was that they could factor in the influence of bed material and consider the effects of using different bed materials. However, the assumption of equilibrium conditions can lead to some deviation from observed ash fusion temperatures. The most decisive manner of determining the ash fusion temperature is thus through bench scale tests.

(3) Influence of feed and bed material composition on melt formation

The alkali index, shown by Equation 2-1, gives the fouling tendencies of biomass fuels based on the K_2O and Na_2O content of the biomass. A threshold limit based on operational experience of 0.34 kg/GJ is suggested (Jenkins *et al.*, 1998); however most biomass fuels exceed this limit by a large extent.

$$AI = \frac{1}{HHV} \cdot \%ash \cdot \%K_2O \cdot \%Na_2O \qquad \textbf{Equation 2-1}$$

where AI = alkali index (kg/GJ)
HHV = higher heating value of the fuel (GJ/kg)
%ash = ash content of fuel
$\%K_2O$ = K_2O content of fuel
$\%Na_2O$ = Na_2O content in fuel

Ohman and Nordin (2000) examined the ash layers on the bed material from a variety of biomass fuels using SEM and found these to be fairly homogenous. The chemical composition of this layer differed significantly to the ash composition of the original fuel

indicating a strong influence of bed material. The deposits on the bed particles could be well represented by the ternary diagram $CaO-K_2O-SiO_2$ and were found mostly in the SiO_2 rich region. The first melting temperature in this region is around 720°C, but an addition of calcium increases this temperature to 1000°C. High potassium content in biomass lowers the ash melting temperature. Coatings with a relatively high fraction of potassium (e.g. samples derived from wheat straw and wood) contained a large amount of melt at temperatures below 900 °C, whereas coatings with a smaller fraction of potassium, (e.g. samples derived from bark, cane trash, and wood residue) did not contain a large amount of melt until temperatures well above 900 °C. Similarly, Zevenhoven-Onderwater (2001) relates the formation of potassium silicate melt to lowering the first melting temperature of silica bed material.

A study done by Sasol aimed at increasing the ash fusion temperature of their dry-bottom coal gasifiers found that adding up to 6% by mass of aluminum oxide to the bed increased the ash fusion temperature of the ash produced in the bed from 1350°C to 1600°C (van Dyk and Waanders, 2007). Other bed materials cited as hindering ash fusion by increasing the ash fusion temperature include kaolin, lime and titanium oxide (van Dyk and Waanders, 2007 and Fryda *et al.*, 2007).

(4) Predicting bed agglomeration temperature for control purposes

An ideal system for preventing bed agglomeration would incorporate an online monitoring system to detect the onset of bed agglomeration. Control measures such as changing the bed temperature, increasing the superficial gas velocity and the bed material make-up rate could then be used to prevent bed agglomeration (Nijenhuis *et al.*, 2007).

Such a system using high frequency pressure drop measurements over the bed was recently developed by Nijenhuis *et al.* (2007) and demonstrated using bench scale fluidised beds. Previously pressure drop over the bed has been used to identify when agglomeration has occurred but not for predictive purposes (Fryda *et al.*, 2007; Ohman and Nordin, 2000). Nijenhuis *et al.* transformed the high frequency pressure drop measurements to an s-statistic using chaos theory. The s-values above 3 indicated a significant change while those below 3 were considered normal. Bench scale agglomeration runs using this technique predicted the bed defluidisation 30 to 45 min before it occurred. Runs to test its control ability were carried out on petroleum coke beds with agglomeration induced by adding water. The s-value rose to

3, and then returned when the bed had stabilised. Tests carried out on an $80MW_{th}$ industrial plant showed that a modified s-statistic could be used successfully.

2.3.4. Summary

Large scale BIG-CCGT units (*i.e.* 100 MW_e) can achieve thermal efficiencies of 45 to 50%. However, the operation of such a large plant is yet to be achieved. Small scale fixed bed gasification systems offer simplicity but efficiencies are similar to combustion systems, *viz.* 30%. Limited experience of biomass gasification in South Africa exists. Notably, Sasol carried out co-firing tests on their fixed-bed, dry-bottom Lurgi gasifier but found that stable operation of the co-fired system resulted in higher CO_2 emissions. Gasification technology for biomass-only feedstreams is not considered to be mature. Tar removal from the fuel gas remains a serious technological hurdle, and in the case of fluidised bed designs, bed agglomeration poses a serious challenge.

2.4. Anaerobic digestion

2.4.1. Process

Anaerobic digestion is the biological conversion of organic material in the absence of oxygen to a.) biogas (which consists of roughly 66% methane and 33% CO_2 (McKendry, 2002b; Wilkie, 2008; Speece, 1996)) b.) additional cell matter and c.) inorganics and refractory material (Wilkie, 2008). The conversion is effected by a complex mixture of bacteria and archae. Anaerobic digestion is a process effective for both wastewater treatment and renewable energy production (Wilkie, 2008). Biogas may be converted to electricity in reciprocating engines or gas turbines or emerging technologies such as micro-turbines or fuel cells. Biogas conditioning may include particulate removal, water knockout and H_2S and ammonia removal (Chambers and Potter, 2002). A comparison of biogas conversion technologies is shown in Table 2-8.

Table 2-8: Comparison of final conversion technologies for biogas (adapted from Chambers and Potter, 2002)

Conversion technology	***Reciprocating gas engine***	***Gas turbine***	***Micro-turbine***	***Fuel cell***
Capacity range (kW)	5 to 50 000	500 to 150 000	30 to 200	<250
Thermal efficiency (%)	20 to 45 %	30 to 50%	25 to 30%	40 to 50%
H_2S tolerance (ppm)	< 200	< 10 000		
Fuel supply pressure	near atmospheric	requires boost compressor	requires boost compressor	
Low LHV tolerance	OK	May require natural gas supplement		
Waste heat conversion	Yes	Yes	No	

The biological conversion process of anaerobic digestion is summarised in Figure 2-12. Complex organics such as polysaccharides, proteins, lipids and nucleic acids are depolymerised through the action of extracellular hydrolytic enzymes to their respective monomers (Schink, 2008). Acidogens break down a variety of carbohydrate substrates to simple organic acids (acetate, butyrate and propionate) hydrogen and carbon dioxide (Flickinger and Drew, 1999; Boone and Mah, 1987). Acetogens, such as *Syntrophobacter* and *Syntrophomonas*, break down propionate and butyrate to acetate and H_2, respectively. Acetotrophic methanogens, such as *Methanosarcina,* convert acetate to methane and carbon dioxide. Hydrogenotrophic methanogens, such as *Methanobacteria,* utilise carbon dioxide and hydrogen to produce methane and water (Waites *et al.*, 2001). Tracer experiments show ~70% of the methane produced in the digester originates from acetotrophs (Ullmans, 2003a).

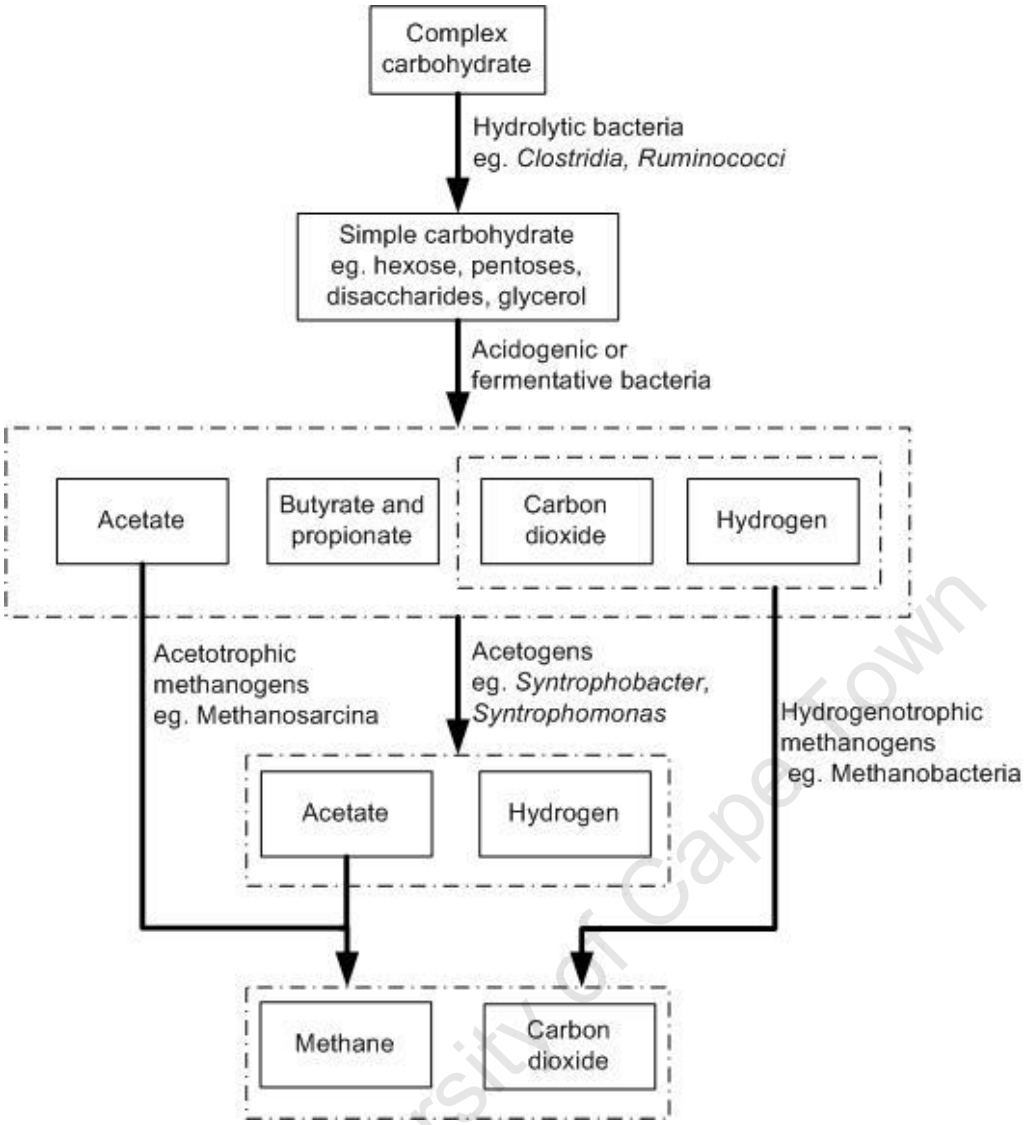


Figure 2-12: Process of anaerobic digestion

Anaerobic digestion cannot proceed if intermediates are allowed to build up. Indeed, the build up of organic acids and concomitant decrease in pH is harmful to the methanogens and their activity drops significantly at pH lower than pH 5 (Speece, 1996). Hydrogen partial pressure (pH_2) plays an important role in regulating the anaerobic digester environment, as shown in Table 2-9. Whilst acetate production is independent of hydrogen partial pressure, the degradation of carbohydrates to butyrate and propionate cannot proceed at high hydrogen partial pressure. Conversely the production of methane from carbon dioxide and hydrogen requires an adequately low hydrogen partial pressure. Thus there is a small "thermodynamic window" in which these reactions can proceed simultaneously (Ullmans, 2003a). By consuming hydrogen, the hydrogenotrophs reduce hydrogen partial pressure and make the conversion of butyrate and propionate to acetate thermodynamically favourable. If the hydrogen is not utilised, it inhibits butyrate and propionate metabolism. Consequently,

accumulation of butyrate and propionate acidify the reactor contents and may lead to the digester failing (Ullmans, 2003a). The interplay of the hydrogen producing acidogens and acetogens and the hydrogen consuming hydrogenotrophs is known as interspecies hydrogen transfer.

Table 2-9: Interspecies hydrogen transfer (adapted form Ullmans, 2003a)

Reaction	***Micro-organism***	***Favourable pH_2***
$CH_3\text{-}COOH \leftrightarrow CH_4 + CO_2$	Acetogenic methanogens	Independent of pH_2
$CH_3\text{-}CH_2\text{-}COOH + 2\ H_2O \leftrightarrow CH_3\text{-}COOH + CO_2 + 3\ H_2$	Acetogens	Low pH_2
$CH_3\text{-}CH_2\text{-}CH_2\text{-}COOH + 2\ H_2O \leftrightarrow 2\ CH_3\text{-}COOH + CO_2 + 2\ H_2$	Acetogens	Low pH_2
$4H_2 + CO_2 \leftrightarrow CH_4 + 2\ H_2O$	Hydrogenotrophic methanogen	High pH_2
$4H_2 + 2\ CO_2 \leftrightarrow CH_3\text{-}COOH + 2\ H_2O$	Hydrogenotrophic methanogen	High pH_2

The minimum generation time is the time a population takes to double in number and is thus inversely indicative of growth rate. Some examples are given in Table 2-10. The long generation time required for acetogens, with concomitant conversion of propionate and butyrate, and the short generation time for the acidogens to produce organic acids means that the build up of acids can easily occur. The increase in pH reduces the activity of the methanogens. The difference in growth rates formed the basis for separating the digestion process into two phases: acid formation phase and methane producing phase. Whilst significant improvements in COD elimination, loading rates and process stability have been noted in such systems, the drawback remains the cost of having separate vessels (Speece, 1996). In an Anaerobic Baffled Reactor (ABR), wastewater moves in plug flow through a series of vertical baffles. This allows the acid forming and methane producing phases to be separated longitudinally down a single reactor (Barber and Stuckey, 1999).The slow growth rate of certain anaerobic digester microbes and the desire to maximise carbon fixation to biogas means that biomass retention in the reactor is important. Increased biomass retention times are achieved in a number of ways: namely, granulation, fixed growth and biomass recycling.

Table 2-10: Minimum generation time for microbes involved in anaerobic digestion (adapted from Speece, 1996)

Substrate	***Micro-organism utilising substrate***	***Minimum generation time under mesophilic conditions (days)***
Simple carbohydrates	Hydrolytic bacteria and acidogenic	0.18
Proteins	Hydrolytic bacteria and acidogenic	0.43
Propionate	Acetogens	3.3
Butyrate	Acetogens	2.0
Acetate	Acetogenotrophic methanogens	3.9
H_2	Hydrogenotrophic methanogens	1.2

Other factors important in the process of anaerobic digestion are temperature and alkalinity. The optimum temperature for mesophilic operations is 35°C and for thermophilic operations 55°C to 65°C. Methanogen activity is more sensitive to temperature change than the acidogens and thus temperature fluctuations can give rise to unstable process conditions (Speece, 1996). Thermophilic operations may achieve metabolic rates that are 50 to 100% higher than mesophilic operations (Speece, 1996). The operation is practised when the reduced reactor size justifies the higher energy requirements and the greater attention to maintaining process stability (Wilke, 2008; Ullmans, 2003a; Tchobanoglous *et al.*, 2003).

The buffering capacity of a digester, as measured by the total alkalinity, is a good indicator of the process stability. The total alkalinity of a well-established digester is 2000 to 5000 mg/l (Tchobanoglous *et al.*, 2003). Buffering substances found in a digester include calcium, magnesium and ammonium bicarbonates and may be supplemented with the addition of caustic soda, lime, sodium bicarbonate or sodium carbonate. Alkalinity is consumed via the generation of acids by the acidogens and the dissolution of CO_2.

2.4.2. Reactor types and current installations

One of the objectives in anaerobic digester reactor design is to separate the biomass retention time or solids retention time (SRT) from the hydraulic retention time (HRT). The HRT defines the time in which the anaerobic microbes can consume the soluble organic load (Speece, 1996). Thus a higher SRT, relative to HRT, allows for greater organic load reductions by retention of a high biomass catalyst concentration. In the case of feedstock containing insoluble complex material, an increased SRT will afford greater opportunity for biodegradation. Reactor types can be defined according to their method of increasing SRT as shown in Table 2-11. Three reactor designs will be considered further due to their applicability in the context of this study, namely, CSTR (Continuous Stirred Tank Reactor); UASB (Upflow Anaerobic Sludge Blanket) and anaerobic lagoons. Small batch anaerobic digesters are not considered viable units for industrial energy recovery as control over process stability is limited (Fannin and Biljetina, 1987). In South Africa, a local company AGAMA has installed a number of batch digesters primarily for household biogas production. Based on data presented in Burton *et al*. (2008) the thermal output of these units is less than 1 kW.

Table 2-11: Types of anaerobic digester reactors

Reactor design	***Method of increasing SRT***
CSTR	SRT = HRT
Upflow Anaerobic Sludge Blanket (UASB)	Microbial granulation
Fixed bed, expanded bed and fluidised bed reactors	Fixed microbial growth
Anaerobic contact process	Biomass recycling
Anaerobic covered lagoon	Plug flow; insoluble solids settle

2.4.2.1. CSTR

In a completely mixed CSTR, the SRT and the HRT are equivalent. However, the design remains popular due to lower initial capital cost and the ability to handle high solids content (Fannin and Biljetina, 1987). The dry solids content of the feed is limited to 7%, after which mixing may become problematic (Fountain, 2008). Hydraulic retention times for a complex substrate such as sewerage sludge range from 15 to 30 days and the volumetric organic load ranges from 1 to 5 kg COD/m^3/day (Tchobanoglous *et al.*, 2003). For simpler substrates, the HRT is less.

Adequate mixing in a CSTR in necessary to prevent the formation of acidified zones. Mixing is achieved through gas injection, stirring or pumped external reticulation. Tank shape has an influence on mixing. Egg-shaped digesters have gained popularity for this reason together with easier sludge withdrawal and better control of the scum layer (Tchobanoglous *et al.*, 2003). The initial cost of construction, however, exceeds cylindrical tank digesters.

A variant of the single stage CSTR design is two stage digestion where a completely mixed digester is followed by a second tank which serves as a settling or holding tank. The design is being phased out due to the capital expense of the second tank with little operational value as less than 10% of total biogas production comes from the second stage (Tchobanoglous *et al.*, 2003). Further, digested solids do not settle well due to entrained gas and fine-sized particles and so the supernatant withdrawn from the second tank may contain high concentrations of suspended solids (Tchobanoglous *et al.*, 2003). Notwithstanding these drawbacks, two-stage digestion still finds application in South Africa, (viz. Cape Town and Johannesburg Wastewater treatment plants (King, 2008 and Deacon, 2008)) illustrating the age of our wastewater treatment plants.

2.4.2.2. UASB

The UASB consists of 3 zones: a granular sludge bed, a sludge blanket and a gas separation and settling zone. Liquid feed is introduced to the reactor via a distributor at the bottom of the reactor. Even distribution is necessary to prevent channelling and bypassing. Some 80 to 90% of the volatile solid destruction occurs in the dense sludge bed section, which occupies 20% of the reactor volume (Fannin and Biljetina, 1987). This results in a large portion of reactor volume which does not contribute to gas production. Gas is collected via internal inverted cones which also serve as settling surfaces to direct solids back into the bed (Fannin and Biljetina, 1987). Upflow velocities in UASB's range from 1 to 3 m/hr (Ullmans, 2003a). This results in reactors with large footprints.

UASB's afford efficient biomass retention without the use of costly packing materials or external solid liquid separation. This allows for high loading rates (Speece, 1996). Good sludge granulation is ssential to the process and this is achieved with soluble feeds. Feeds with high suspended solids inhibit the formation of granules (Tchobanoglous *et al.*, 2003). Whilst the properties of the microbial granules in the dense sludge bed are central to the process, their formation is still largely empirical (Ullmans, 2003a). Indeed, to start up a UASB, sludge is generally imported from an operational unit to reduce start up time (Tchobanoglous *et al.*, 2003).

UASB reactors are considered to be a proven process and more than 500 large scale units exist (Ullmans, 2003a). One of the earliest installations was the South African "anaerobic clarigester" for treating winery wastewater (Speece, 1996). UASB's have been used to treat sugar wastewaters successfully (Speece, 1996) as well as sugar beet and potato starch wastewaters (Biljetina, 1987). SAB Durban brewery has used UASB technology since 1996 (Speece, 1996) and SAB Newlands Brewery has treated their wastewater with a UASB reactor since 2002 (Hoffman, 2008). The use of this technology has subsequently been extended to 5 other breweries. One of the original drivers for installing the UASB at Newlands was to mitigate the cost of effluent discharge in the municipal waste water system (Hofmann, 2008).

2.4.2.3. Covered lagoons

Covered lagoons operate in plug flow with the wastewater introduced at one end of the lagoon and withdrawn at the other. This digester type has been used to treat manure from animal husbandry and can also handle fats and grease. HRT ranges from 30-50 days and it is estimated that the SRT ranges from 50-100 days. Baffles are often installed to prevent the solids washing out before digestion has occurred. Care must be taken to remove non-biodegradable material (Tchobanoglous *et al.*, 2003).

2.4.3. Challenges

Whilst significant energy recovery is possible from anaerobic digesters, it is important to bear in mind that the primary driver in most installations has been wastewater treatment. As such, energy recovery systems have typically not been installed and it is common to find biogas being flared. Exceptions in SA include the PetroSA wastewater treatment plant, where the biogas is currently being used to generate electricity. This project is a registered CDM (Carbon Development Mechanism) project and hence was only economically feasible with the sale of Certified Emission Reductions (CER's or "carbon credits"). In some cases the economic benefits of energy recovery from biogas is being recognised by industry. However, this requires adjustment of the process control philosophy. For example, the SAB Newlands process control philosophy aims to treat the maximum amount of effluent at the expense of a fluctuating biogas flow. To be able to run a biogas burner, the flow needs to be relatively constant (Hofmann, 2008).

2.4.4. Feedstock pretreatment

Certain components in biogenic waste streams (e.g. cellulose and microbial cell walls) are resistant to biological action and unless the feed is pretreated in some manner, these substances pass unconverted through the system. Various methods of pretreating the feed exist to make it more amenable to hydrolysis. Work done on pre-treating lignocellulosic feedstocks for fermentation to ethanol (Sun and Cheng, 2002; Wyman *et al.*, 2005) applies to pretreating feed for anaerobic digestion as both processes require hydrolysis of the feed before fermentation or digestion can take place. A number of methods of disrupting the cell wall to release the intracellular proteins, lipids, nucleic acids and carbohydrates exist *e.g.* hot water hydrolysis (with and without chemical additions), high pressure homogenisation and sonication. Hot water hydrolysis and high pressure homogenisation already exist as patented

processes for pretreating sewerage sludge (Kelly, 2006 and Stephenson *et al.*, 2007). Sonication has yet to gain industrial application, primarily due to the high energy inputs required (Weemaes and Verstraete, 1998).

2.4.4.1. Pretreatments for lignocellulosic waste

Lignocellulosic treatments reviewed here are: dilute acid treatment, hot water hydrolysis, steam hydrolysis, ammonia pretreatment and lime pretreatment. Process conditions and modes of action of each pretreatment are summarised in Table 2-13. Alkali pretreatments are excluded from the analysis owing to the reagent cost and waste processing (Wyman *et al.*, 2005a).

Physical factors which affect the enzymatic hydrolysis of lignocellulosic material are: cellulose porosity, cellulose fiber crystallinity, lignin content and hemicellulose content. Cellulose porosity is a measure of surface area accessible to enzymes. Cellulose fiber crystallinity refers to the ratio of amorphous to crystalline cellulose. A pretreatment process will thus seek to increase porosity, reduce cellulose crystallinity, remove hemicellulose and lignin from the microfibril matrix or alter the lignin structure to expose the microfibril matrix (Sun and Cheng, 2002). Sun and Cheng identify four requirements for successful pretreatment:

- Optimal formation of sugars or the ability for subsequent sugar formation by enzymatic hydrolysis
- Minimal degradation or loss of carbohydrate
- Absence of byproducts that inhibit the subsequent hydrolysis and fermentation processes
- Cost-effectiveness

A mechanical pretreatment is usually employed upstream of the pretreatment to produce a homogenous feedstock of small particles. The energy consumption of knife and hammer mills is shown in Figure 2-13. It can be seen a feedstocks with high lignin content (*e.g.* hardwood) require greater energy input than those with lower lignin content (*e.g.* corn stover).

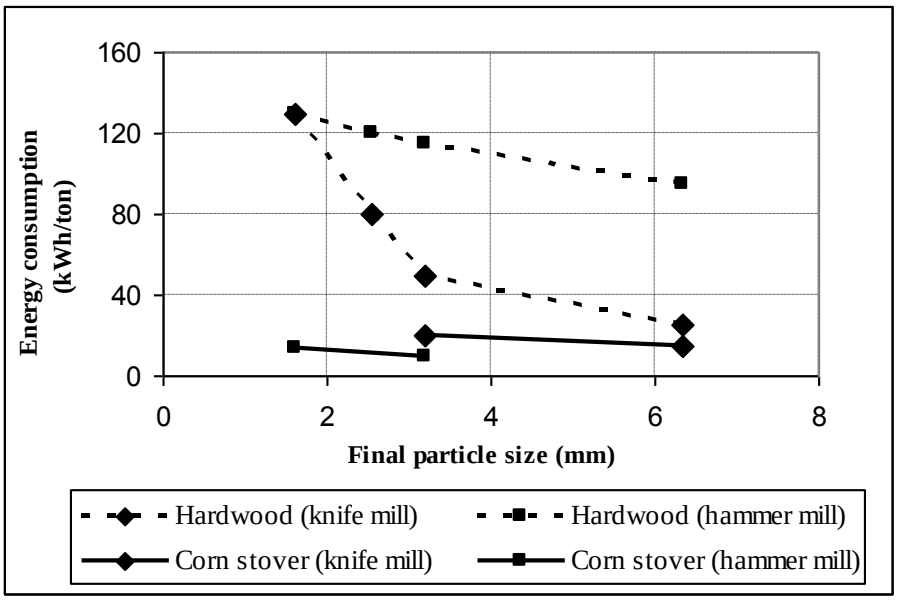


Figure 2-13: Energy consumption of particle size reduction techniques (adapted from Sun and Cheng, 2002)

(1) Dilute acid pretreatments

A number of variations of dilute acid pretreatments exist. In the one model, 0.5-3.0% sulphuric acid is fed co-currently with biomass to a high pressure reactor (Wyman et al, 2005). It is proposed the acid acts as a catalyst to cellulose conversion to glucose. The lignin is disrupted, increasing cellulose susceptibility to enzyme action. This method has also been used in the production of furfural from biomass. In the percolation model, a dilute solution of sulphuric acid is passed through the biomass (Wyman et al, 2005). Whilst the method displays almost 100% hemicellulose removal and a subsequent 90% enzymatic conversion of cellulose, its water requirement is too high for commercial application (Mosier, 2005). The use of sulphur dioxide yields similar results and without the attendant water requirements, but is expensive. In dilute acid processes, non-productive binding of enzymes to remaining lignin also occurs and the treated liquor needs to be neutralized and conditioned prior to the biological steps.

(2) Hot water hydrolysis and steam explosion

Hot water hydrolysis works in the same manner as dilute acid pretreatments, as the pH of water at the high temperatures and pressures used in the process is low. It is considered preferable to acid pretreatments as no extraneous reagent or downstream neutralisation is needed. Up to 22% of the cellulose, 35 to 60% of the lignin and all of the hemicellulose is removed from the fiber matrix. This renders the cellulose more susceptible to enzyme action.

Over 90% of the hemicellulose is recovered as monosaccharides when acid is used to hydrolyse the resulting liquid (Mosier *et al.*, 2005), although the monosaccharides formed in the reaction step can be degraded (Wyman *et al.*, 2005).

Steam explosion involves treating coarsely chipped biomass with high-pressure saturated steam, followed by rapid pressure reduction to make the materials undergo an explosive decompression (Sun and Cheng, 2002). A large energy saving is realised in avoiding communition to fine particles as larger particles (5 cm diameter) respond well to this treatment (Cullis *et al*., 2004).

(3) Ammonia pretreatments

Ammonia pretreatments include: Ammonia Fibre EXplosion (AFEX), Ammonia Recycle Percolation (ARP) and Soaking Aqueous Ammonia (SAA). In AFEX the feedstock is exposed to liquid ammonia at high temperature and pressure. The pressure is then suddenly released and the material undergoes explosive decompression; the ammonia volatises and can be recycled (Sun and Cheng, 2002). Very little liquor and soluble products are produced in the process but it opens the lignin structure and depolymerises the hemicellulose to oligomers. It is believed that ammonolysis of glucuronic cross-links makes the carbohydrate more accessible to enzymatic hydrolysis (Wyman *et al.*, 2005b). Whilst oligomers cannot be fermented to ethanol, they can be digested by digester fermentative bacteria (Tsao, 1987). The ammonia is not harmful to downstream micro-organisms and can be used on coarse biomass particles (Wymans *et al*., 2005a). The process has not demonstrated success with high lignin containing biomass; the enzymatic cellulose conversion fell from 90% to 50% when the lignin content increased from 5 to 20% (Sun and Cheng, 2002). In ARP aqueous ammonia is recycled through the feedstock at elevated temperatures. SAA is a simpler batch process, where the biomass is soaked in an aqueous ammonia solution at ambient conditions.

(4) Lime pretreatments

Low temperature lime pretreatment is a simple, inexpensive option. Operating temperatures range from 25 to 130 °C, and the corresponding treatment time ranges from weeks (25 °C) to hours (130 °C) (Wyman *et al.*, 2005a). The process removes approximately 33% of lignin (in case of low-lignin content herbaceous materials) and 100% of acetyl groups.

2.4.4.2. Comparison of lignocellulosic pretreatments

To obtain comparative xylose and glucose yields, Wyman et al (2005b) carried out a co-ordinated inter-laboratory study using a single source of corn stover, the same cellulase enzyme, shared analytical methods and common data interpretation approaches. Sugar yields are shown for the pretreatment (Stage 1) and for the hydrolysate treated by 15 FPU/g glucan of cellulase enzymes (Stage 2). Results are expressed as a percentage of the theoretical sugar yield and are presented in Table 2-12. The results show that the pretreatment themselves result in less sugar release than the enzyme treatment; it is the action of opening up the fibre matrix which allows the enzymes to act effectively.

Table 2-12: Comparison of xylose and glucose yields for selected lignocellulosic pretreatments

	Percent of theoretical sugar yield by enzymatic hydrolysis		
Pretreatment system	***Stage 1***	***Stage 2***	***Total***
Dilute acid (co-current)	36%	56%	92%
Dilute acid (percolation)	41%	56%	97%
Hot water hydrolysis	25%	62%	87%
AFEX	0%	94%	94%
ARP	18%	72%	89%
Lime	10%	77%	87%

Table 2-13: Summary of operating conditions for pretreatment technologies

Pretreatment technology	Chemicals used	Batch or continuous process	Temperature (°C)	Pressure (atm)	Reaction time (min)	Solids concentration (wt %)	Mode of action [c]
Dilute sulphuric acid (co current) [a]	0.5–3.0% sulphuric acid	Continuous	130–200	3–15	2–30	10–40	Increases accessible surface area; removes hemicellulose; alters lignin structure
Dilute sulphuric acid (percolation) [a]	0.07–0.1% sulphuric acid	Continuous	190–200	20–24	12–24	2–4	Increases accessible surface area; removes hemicellulose; alters lignin structure; removes lignin
Hot water hydrolysis [a]	Water	Semi-batch	160–190	6–14	10–30	5–30	Increases accessible surface area; removes hemicellulose; alters lignin structure
Steam explosion [b]	Steam	Batch	160–260	6.8 – 48	1 – 10	33	Increases accessible surface area; removes hemicellulose; alters lignin structure
AFEX [d]	1kg liquid NH_3/kg biomass	Semi-batch	70–90	15–20	<5	60–90	Increases accessible surface area; decrystallises cellulose; removes lignin; alters lignin structure; removes hemicellulose
ARP [a]	10–15 wt.% ammonia	Continuous	150–170	9–17	10–20	15–30	Increases accessible surface area; decrystallises cellulose; removes lignin; alters lignin structure; removes hemicellulose
SAA [a]	Dilute aqueous ammonia	Batch	25–60	1			
Lime [a]	0.05–0.15 g Ca(OH)2/g biomass	Batch	70–130	1–6	1–6 h	5–20	Increases accessible surface area; removes lignin; alters lignin structure; removes hemicellulose
Lime + air [a]	0.05–0.15 g Ca(OH)2/g biomass	Batch	25–60	1	2 weeks–2 months	10–20	

a.) Wyman et al, 2005a
b.) Sun and Cheng, 2002
c.) Mosier *et al.*, 2005
d.) Teymouri et al, 2005

2.4.4.3. Pretreatment methods for microbial sludges

(1) Hot water hydrolysis

The initial goal of hot water hydrolysis was to improve the dewaterability of sludge (Weemaes and Verstraete, 1998) as the breakdown of the cell wall releases the cell bound water. However, its application to improve digestibility and concomitant biogas production has been recognised. The patented CAMBI^{TM} process has been employed successfully in a number of European cities (Kelly, 2006). The CAMBI^{TM} process runs at 165°C and a pressure of 8 bar for 30 min and does not require the addition of chemicals. Anaerobic digestion of the treated sludge generally results in an increase from 30% volatile solids reduction to 60% (Fountain, 2008). Work by Kim *et al.* (2003) showed that the addition of 7g/l NaOH to a 121°C, 30 min hot water hydrolysis process, resulted in a COD solubilisation of 80%. The addition of acid also improves COD solubilisation (Neyens and Baeyens, 2003). However, the hydrolyzate needs to be neutralised before entering the digester (Valo *et al.*, 2004).

The operation of the CAMBI^{TM} thermal hydrolysis pretreatment process has been described in section 3.3.3. The sludge is pumped into a high pressure "pulper" vessel where it is raised to near its saturation temperature by the addition of steam. The sludge is held at 165°C and 8 bar for 30 min. Literature cites the temperature range for thermal hydrolysis processes as 60 to 180°C (Neyens and Baeyens, 2003). Temperatures greater than 150°C showed marked increase in cell disruption, though temperatures greater than 180°C result in the formation of refractory compounds. Holding times ranged from 10 to 30 min, with longer holding times not showing marked increases in cell disruption. Sulphuric acid and bases (KOH, NaOH, $Ca(OH)_2$) may be added to increase cell disruption.

(2) High pressure homogenisation

High pressure homogenisation is adapted from the food and pharmaceutical industry, where it is used to create emulsions, to use in cell disruption in the biotechnology field to obtain intracellular products. It is the preferred method for disrupting non-filamentous cells (Harrison 1991; Middelberg, 1995). A positive displacement pump in small scale units or a multiple piston pump in large-scale units provides a pressurised cell suspension at 20 to 120 MPa (Middelberg, 1995). The pressurised cell suspension is forced through the centre of a

valve seat, then radially across the seat face before striking an impact ring. The valve aperture ranges from 10 to 20µm and radial velocities can reach to 300m/s.

Gram-negative bacteria were found to be easier to disrupt than Gram-positive bacteria; requiring a 60% increase in pressure to achieve the same amount of cell disruption. Operating pressures for the disruption of bacteria for protein release ranged from 15 to 24 MPa and biomass concentrations ranged from 95 to 260 kg dry solids/m^3 (Harrison, 1995). The specific energy requirement for high pressure homogenisation of gram-negative cells is reported to be 0.8 to 1.2 J. The percentage disruption achieved was 90 to 87% (Engler, 1985). Energy requirements for laboratory batch operations may be calculated from Equation 2-2 (Anand et al, 2007).

$$E = Q \cdot \Delta P \cdot N$$ **Equation 2-2**

where E = energy consumption (J)
ΔP = operating pressure of unit (Pa)
Q = volumetric flowrate (m^3/s)
N = number of passes

The patented Microsludge process, which makes use of high pressure homogenisation, has been employed in two American wastewater treatment plants (Stephenson *et al.*, 2007). Operating conditions such as pressure and number of passes are not given in the literature. In other experimental work on the disruption of waste activate sludge; operating pressures of 30 MPa, (Nah *et al.*); 50 MPa (Onyeche and Schafer, 2003) and 60 MPa (Weemaes and Verstaete, 1998) are cited. The Microsludge process involves passing thickened secondary activated sludge through a coarse filter, after which NaOH is added. The suspension is mixed by a high shear mixer before being stored in conditioning tank. The conditioned sludge is then transferred to a gas liquid seperator and through a fine filter before entering the high pressure homogeniser. The process achieves 80% volatile solids reduction in some cases (Stephenson, 2006), but it would appear that this is variable (Stephenson *et al.*, 2007). The increase of temperature due to the dissipation of viscous energy benefits the process as it negates the need for feed preheating.

2.4.5. Summary

Anaerobic digesters are primarily installed for wastewater treatment purposes and the technology is well-established in this field. However, the advantages of using the biogas from anaerobic digestion for plant-wide heat and power are being recognised. A wide range of reactor designs exist; CSTR's and covered lagoons offer the simplest designs and high rate

digesters such as UASB's offer higher throughput. There is considerable operational experience of CSTR and UASB designs in South Africa, particularly at municipal wastewater treatment plants and breweries, respectively. The application of anaerobic digestion to lignocellulosic feedstocks requires some form of pretreatment to aid the digester hydrolytic bacteria. Microbial sludges feedstocks also benefit from pretreatment.

2.5. Conclusion

Of the three technologies reviewed here, combustion and anaerobic digestion offer the most readily implementable means to process biogenic wastes to heat and electricity in South Africa. This is due primarily to technology maturity and the operational experience which exists in South Africa. Biomass gasification requires further advancements in gas cleaning technology if the gas is to be used in an internal combustion engine or gas turbine. Preliminary tests at the Sasol Secunda gasifier have shown that even co-firing of biomass is not without its drawbacks. For example, stable operation was achieved at higher steam:oxygen ratios which resulted in higher CO_2 production. Furthermore, immature gas cleaning technology, high cost of delivered biomass and low fixed-carbon of biomass were cited as reasons for not pursuing the project.

Chapter 3 Comparison of combustion and anaerobic digestion for processing biogenic waste – a case study

3.1. Introduction

In this chapter, the potential of combustion and anaerobic digestion for energy generation from sewerage sludge is investigated through case studies of three Thames Water sewerage treatment plants. Strict legislation in the UK regarding the disposal of sewerage sludge to land has resulted in the wastewater treatment plants processing their sludge further via combustion or anaerobic digestion to minimise disposal costs and simultaneously generate renewable electricity. Similar data is not available in South Africa as energy recovery from anaerobic digestion and combustion is not commonly practised on our wastewater treatment plants.

3.1.1. Thames Water company overview

Thames Water is a UK-based utilities company that supplies clean water to London and the Southeast of England. In addition to water supply, Thames Water treats domestic and industrial wastewater in 350 water treatment plants. Thames Water was originally formed to oversee a number of smaller water boards and to regulate the Thames catchment area. It was privatised in 1989 and the regulatory responsibilities were taken up by government. In 2001 the company was acquired by a German utility, RWE. In 2006 RWE sold Thames Water to Kemble Water, a consortium managed by Macquarie Capital Funds (thameswater.co.uk, 2008). Under the new ownership an investment of £1 billion was made in 2007/2008. Upcoming projects include the construction of a desalination plant in Beckton, London and construction of tideway tunnels to prevent sewerage outfall into the Thames.

3.1.2. EU and UK legislature regarding sewerage sludge disposal

The UK adopted the EC directive on urban wastewater treatment in 1994 (OPSI, 1994). Included in this legislature was the banning of sewerage sludge disposal to sea. The EC directive on sewerage sludge seeks to promote the use of treated sewerage sludge in agricultural land application provided it complies with applicable requirements of monitoring, treatment and quality (ec.europa.eu, 2001). In the UK, the application of untreated sewage sludge to land used to grow food crops ceased at the end of 1999. The application of untreated sewage on land used to grow industrial crops that may also have had a food use was phased

out at the end of 2001 and the application of untreated sewage sludge on agricultural land used to grow non-food crops ceased with effect from 31st December 2005 (genesisqa.com, 2008). This legislation resulted in an increased cost of sewerage sludge disposal as essentially all land disposal options required some form of sludge treatment. Wastewater treatment plants have thus reviewed their sludge disposal options in terms of benefits gained from the various treatment options. Thames Water makes use of four sludge disposal methods:

1. Treating the sludge with lime to kill pathogens prior to its application to agricultural land
2. Anaerobically digesting the sludge, followed by application of the digested sludge to land.
3. Pasteurising the sludge prior to digestion to increase its range of application, followed by application of the digested sludge to land
4. Using enhanced anaerobic digestion (*i.e.* digestion with some form of pretreatment) to effect greater volatile solids destruction and concomitant volume reduction, followed by application of the digested sludge to land
5. Incineration (combustion) of the sludge and disposal of the ash to landfill.

The latter four options have potential for energy recovery through electricity generation or raising process steam either by combined heat and power engines or steam cycle turbines.

3.1.3. Objectives

A number of visits to Thames Water treatment plants were made between November 2007 and February 2008. The plant visits focused on the integration of sludge treatment and energy generation. The technologies considered are: a.) anaerobic digestion, b.) enhanced anaerobic digestion and c.) combustion. Much insight into operational and maintenance issues was obtained from these visits and the presentation of the field notes from these trips is included in this chapter with a view to exposing the South Africa audience to the energy generation potential of wastewater treatment plants and associated difficulties. Furthermore, the objectives of this study are to:

- Develop energy yield models for combustion and anaerobic digestion
- Compare the predicted results to plant data
- Determine which technology gives the highest energy recovery

3.2. Overview of wastewater treatment process

3.2.1. Wastewater treatment process

The main steps in treating the wastewater are: 1.) fat, oil, grit and grease (FOGG) removal, 2.) primary settling, 3.) biological treatment and 4.) secondary settling. Wastewater is collected via sewers and pumped to the wastewater treatment plant. Coarse (6 to 24mm) and fine (2 to 6mm) screens at the inflow to the wastewater treatment plant (WWTP) remove large floatable debris which are disposed of at landfill (Waites *et al.*, 2001). Degritting channels slow the flow down to allow grit and other remaining solids to settle out. These solids are disposed of at landfill or used as building material (Waites et al, 2001). The flow may then pass through comminutors or macerators to break up any remaining solids to fine particles. Rag (toilet paper) is particularly problematic; if not removed in screening step it can clog the macerators and is not easily settled in the primary settlers and not easily degraded in the biological treatment (Fountain, 2008).

While both aerobic and anaerobic treatments of the the wastewater are applicable, aerobic treatment has found preference owing to lower concentrations of biodegradable COD, higher effluent quality needs and nutrient removal requirements (Tchobanoglous *et al.*, 2003). Two classes of aerobic biological treatments will be considered here: a suspended growth process (activated sludge process) and an attached growth process (trickling bed filters).

Prior to aerobic biological treatment, 50 to 70% of suspended solids are removed in primary thickeners, giving a 30 to 40% reduction in BOD (Waites *et al.*, 2001). The underflow from the primary settlers is called primary sludge. The overflow is sent to the biological treatment plant where soluble, colloidal and suspended biodegradable material is removed (Tchobanoglous *et al.*, 2003).

In an activated sludge process, vigorous mixing through aeration allows the aerobic bacteria to metabolise the biodegradable material. Depending on the micro-organism to food ratio, two modes of metabolism occur:

1. Oxidation and synthesis

   $C_aO_bH_cN_dS_e + O_2 + \text{nutrients} \rightarrow CO_2 + NH_3 + C_5H_7NO_2 + \text{other end products}$

2. Endogenous respiration

   $C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3 + energy$

It is desirable to operate at high micro-organism to food ratio to encourage endogenous respiration and decrease the sludge volume that must be handled. A decrease in sludge volume is also achieved by operating at high SRT (Tchobanoglous *et al.*, 2003).

An activated sludge process generally incorporates aerobic as well as anoxic (absence of oxygen) zones where nitrifying and denitrifying bacteria remove nitrogenous waste by converting ammonia to nitrate and nitrate to nitrogen respectively. The overflow from the secondary settlers is generally safe to be released to the environment although some plants may include tertiary treatment or "polishing". A portion of the underflow from the secondary thickeners is returned to the beginning of the process (returned activated sludge (RAS)) to maintain the microbial culture density. The excess or waste activated sludge (WAS) is sent, along with the primary sludge, to be dewatered and further treated.

Trickle bed filters are generally circular, sunken tanks filled with a packing on which the microbial population grows. Pebbles may be used as packing and have the advantage of being robust but cheaper plastic packing is becoming more common. The water is trickled over the surface of the vessel by a rotating feed arm. Sludge arises when the microbial colonies slough off the packing. This is known as humus sludge and significantly smaller amounts are produced compared to activated sludge process. It is also more amenable to hydrolysis. The humus sludge is settled in secondary thickeners from where the overflow is released to the environment and the underflow is sent to be dewatered and further treated.

The underflows from the primary and secondary settling tanks, containing 2% and 1% solids respectively, are dewatered to around 6% solids, typically using gravity belt thickeners. Secondary sludge does not dewater as well as primary sludge due to the intra-cellular water of the bacteria contained in the secondary sludge. Free water (the water in the interstitial spaces) can be removed in a dewatering process whereas intracellular water, comprising some 70% of microbial biomass, cannot. In general, fixed film biosolids (*i.e.* humus sludge) have a higher specific gravity than WAS and hence higher settling rates which allows for better dewatering.

3.2.2. Sludge treatment options

Owing to the pathogenic nature of the sludge generated in the primary and secondary settlers, further treatment is needed before disposal. Treatment options prior to final disposal to land are reviewed below and opportunities for energy recovery are highlighted.

3.2.2.1. Alkaline stabilisation

Alkaline stabilisation requires the addition of hydrated lime ($Ca(OH)_2$) or quicklime (CaO) in sufficient quantity to raise the pH to 12 or higher. This halts the microbial reactions responsible for odour formation and vector attraction (Tchobanoglous *et al.*, 2003). If quicklime is added the hydration reaction releases enough heat to raise the temperature to 50°C which inactivates worm eggs (Tchobanoglous *et al.*, 2003). Hydrated lime can be applied prior to dewatering, but a number of disadvantages are associated with this method including: scaling problems in dewatering equipment and as in the case of using hydrated lime, larger volumes have to be processed. Dosage rates range from 120 g $Ca(OH)_2$/kg dry solids (primary sludge) to 300 g $Ca(OH)_2$/kg dry solids (waste activated sludge). Quicklime can be added dry to dewatered sludge. The higher the solids content of the sludge and the greater the loading rate, the higher the temperature increase, e.g. adding 300 g CaO/kg dry solids to sludge with a solids content of 20% results in a temperature rise of 18°C; adding the same amount to 16% dry solids sludge increases the temperature by 15°C (Tchobanoglous *et al.*, 2003). No energy recovery is possible from alkaline stabilisation.

3.2.2.2. Composting

Composting produces a nuisance-free, humus-like material which can be applied to land. The sludge undergoes aerobic degradation converting 20 to 30% of the volatile solids to water and carbon dioxide. The process is aided by aeration; using either a blower system under the pile or by regular turning of the heap or windrow. Heat release during degradation can result in compost heap temperatures of 50 to 70°C, allowing pasteurisation to occur (Tchobanoglous *et al.*, 2003). Goudkoppies WWTP in Johannesburg trialled a composting process, (illustrated in Figure 3-1), in 2006 with some success (Boyd, 2008). The resulting Class A sludge has an Na-P-K composition given in Table 3-1, and can be sold as a sludge fertiliser. Class A sludges must contain less than 1000 CFU of E. Coli per g dry solid and less than 0.25 viable *Ascaris* ova per g dry solid as laid out in the 2006 guidelines for application of sludge to land (Snyman and Herselman, 2006b). Composting does not allow for energy recovery, the

process is in fact energy consuming. However, it does play an important role in returning nutrients to the soil.

Table 3-1: Properties of Goudkoppies compost (Boyd, 2008)

Parameter	*Units*	*Quantity*
Kjeldhal nitrogen as N	g/kg dry solids	10
Total phosphorus as P	g/kg dry solids	26
Potassium as K	g/kg dry solids	2
Total dry solids	%	53%
F Coliform	CFU/g dry solids	Not detectable
Viable *Ascaris* ova	ova/g dry solids	Not detectable

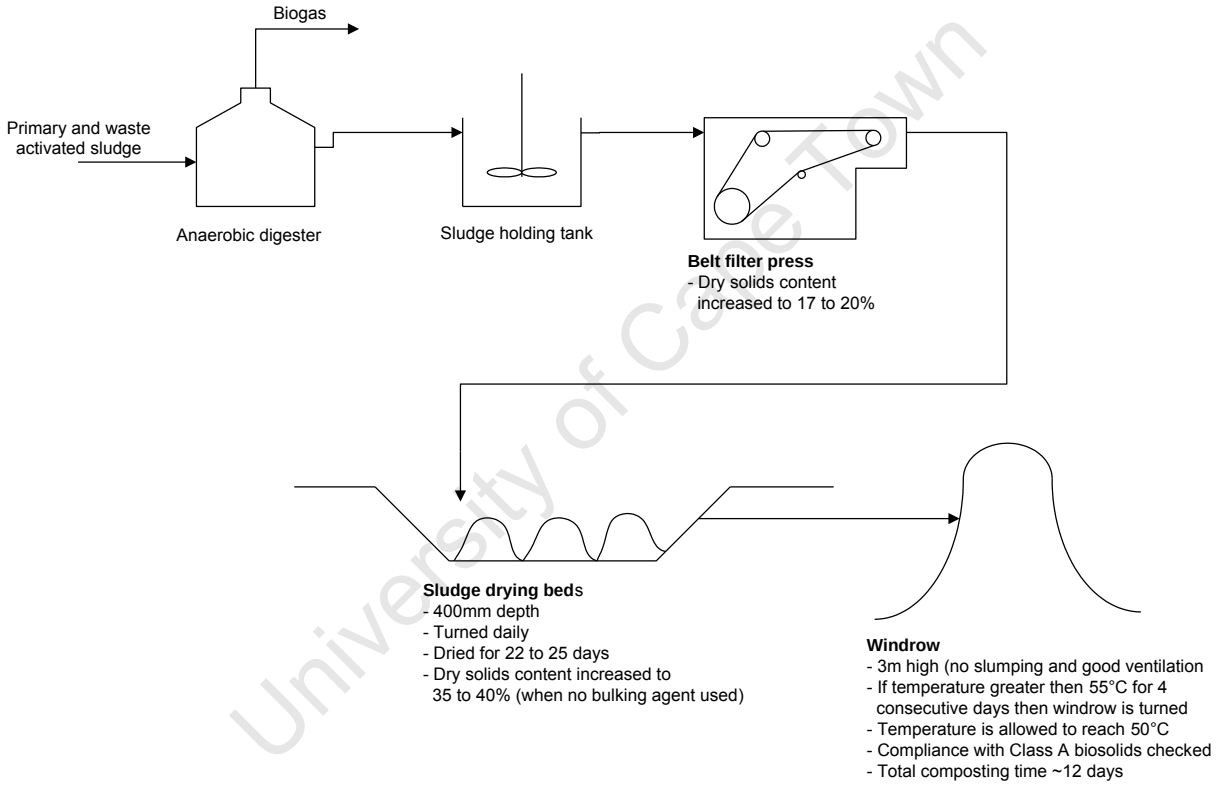


Figure 3-1: Composting process used at Goudkoppies WWTP, Johannesburg (Boyd, 2008)

3.2.2.3. Aerobic digestion

Aerobic digestion proceeds in much the same way as the activated sludge process. To reduce the amount of new cells produced, the process is run under substrate limitation to force the cells to respire endogenously. This results in an overall reduction of volatile solids. Aerobic digestion requires the use of energy intensive compressors for aeration. The resulting sludge needs final disposal to land. Aerobic treatment is employed as a sludge stabilisation process on 20% of the plants surveyed by Snyman *et al.* (2004).

3.2.2.4. Anaerobic digestion

Anaerobic digestion is the oldest form of sludge stabilisation. The process has been described in Section 2.3. Anaerobic digestion has two advantages over aerobic digestion: a.) the production of biogas and b.) a greater reduction in volatile solids. Digested sludge should have a low odour and humus-like quality.

At Thames Water, volatile solids destruction ranges from 65% in processes which have pretreatments, to 35-45% for those that do not. The volatile solids destruction achieved in South African WWTP's are given in Table 3-2. Anaerobic digestion is the most widely used sludge stabilisation method in South Africa and is employed in 57% of plants surveyed by Snyman *et al.* (2004).

Table 3-2: Volatile solids destruction achieved in South African municipal WWTP's

Metropole	% volatile solids destruction	Comment	Reference
Johannesburg	25%		Deacon, 2008
Cape Town	10-15%	Calculated from Cape Flats and Athlone WWTP	King, 2008
Durban	50-70%	Seems implausibly high	Dildar, 2008
Tshwane	49%	Design figure only	Ntsowe, 2008

3.2.2.5. Incineration

Bubbling fluidised bed combustors and other designs have been reviewed in Section 2.2. Multiple hearth furnaces and bubbling fluidised beds may be used to incinerate sludge, with the latter becoming increasing more popular. In Germany, 27 fluidised bed combustors are used for sewerage sludge incineration. In the USA, the figure is over 100 units built (Werther and Ogada, 1999). In the UK, 20% of the sludge generated is incinerated (Cheeseman *et al*., 2003). In South Africa no sludge is currently incinerated though the eThekwini municipality does have an incinerator installed (Dildar, 2008). The ash generated from incineration is generally sent to landfill. Depending on the heavy metal content, the ash may be used as an agricultural fertiliser or as a clay-replacement agent in brick-making (Cheeseman *et al*., 2003).

3.2.3. *Final disposal of sludge to agricultural land*

Guidelines regarding the application of sewerage sludge to agricultural land in South Africa are given by the WRC (Snyman and Herselman, 2006a). The 1997 classification system and associated disposal options are shown in Table 3-3. In their comprehensive survey of South African WWTP's, Snyman *et al*., (2004) found that 24% of sludge was Class A, 51% was Class B, 14% was Class C and 11% was Class D according to the 1997 classifications. Snyman *et al.* go on to comment that at many of the WWTP's surveyed the sludge was stockpiled in unlined dams.

Table 3-3: 1997 WRC sludge classification system

Type of sewage sludge	Examples	Characteristics and quality of sludge
TYPE A	Raw sludge; cold digested sludge, septic tank sludge, oxidation pond sludge	• Usually unstable and can cause odour nuisances and fly-breeding
		• Contains pathogenic organisms
		• Variable metal and inorganic content
TYPE B	Anaerobic digested sludge (heated digester); WAS, humus sludge	• Fully or partially stabilised - should not cause significant odour nuisance or fly-breeding
		• Contains pathogenic organisms
		• Variable metal and inorganic content
TYPE C	Pasteurised sludge; heat-treated sludge, lime-stabilised sludge; composted sludge; irradiated sludge	• Certified to comply with the following quality requirements
		1.) Stabilised (should not cause odour nuisances or fly-breeding)
		2.) Contains no viable *Ascaris* ova per 10 g dry sludge
		3.) Maximum 0 *Salmonella* organisms per 10 g dry sludge
		4.) Maximum 1000 Faecal coliform per 10 g dry sludge, immediately after treatment (disinfection/sterilisation)
		• Variable metal and inorganic content
TYPE D	Pasteurised sludge; heat-treated sludge, lime-stabilised sludge; composted sludge; irradiated sludge	• Certified to comply with the same quality requirements as the TYPE C sludge.
		• Limit on metal and inorganic content dry sludge
		• User must be informed about the moisture and N P K content.
		• User must be warned that not more than 8 t ha^{-1} yr^{-1} (or 0.8 kg/m^2) (dry sludge) may be applied to soil and that the pH of the soil should preferably be higher than 6.5.

3.3. Thames Water' plants investigated

Three Thames Water wastewater treatment plants in and around London, UK were visited. The process description, process block diagram and plant operation are described in this section.

3.3.1. Beckton WWTP, London Docklands

3.3.1.1. Process description

The Beckton wastewater treatment plant (WWTP) located in London Docklands is the largest in the UK. The plant employs the activated sludge process and incineration is the final sludge treatment option. Figure 3-3 shows an aerial photograph of Beckton WWTP.

The process at Beckton WWTP is a typical activated sludge process. Refering to Figure 3-2: water from sewerage lines is channelled through the degritting channels to remove "rag" (toilet paper) and other large floatable material. The flow is then slowed down, and any sand from storm water run-off is allowed to settle. The water then passes to the primary settling tanks, which are rectangular vessels with top skimmers to remove oil and grease and a slanted bottom to bring the sludge to a central collection drain. The underflow is thickened on primary gravity belt thickeners and the overflow enters the activated sludge process. The activated sludge is thickened to 1 to 2% dry solids content in the secondary thickeners. The overflow from the secondary settlers is returned to the river. Part of the thickened sludge in the underflow is recycled back (the Return Activated Sludge or RAS) to the top of the activated sludge process to maintain the culture in the vessel, and the rest (the Surplus Activated Sludge or SAS) is sent to the secondary gravity belt thickeners. At Beckton WWTP the ratio of RAS to SAS is 9:1.

Primary and secondary thickened sludge is mixed prior to the sludge holding tanks and is then pumped to a filter press. The filter press operates up to 10 bar producing a filter cake of ~20% solids conten which is conveyed to the paddle wheel feeder of the fluidised bed incinerator. The bed of sand is fluidised by air passing through the distributor plate and the temperature is maintained at 750°C by the combustion of the sludge. Should the temperature drop due to bad combustion conditions (*e.g.* cake too moist); auxiliary gas lances are fired into the bed. The flue gases leave the bed at 850 to 900°C. The hot gas is used to raise steam for a steam turbine

and is then passed through a heat exchanger to preheat the incoming air stream to 300°C. The exhausted steam from the turbine is sent to an economiser to further energy recovery. The cool flue gases pass through bag filters to remove fine particulate matter before being released to the atmosphere. Ash is collected from the bed and is trucked to a landfill site. shows the block flow diagram of the Beckton WWTP.

3.3.1.2. Challenges to plant operation

Since commissioning of the fluidised bed incinerator there have been two serious occurrences of bed agglomeration, which in plant shut-downs, removal of agglomerated material and refurbishment of the incinerator.

The sequence of events which led to the second incident are:

- The overbed feeder system did not distribute the fuel adequately over the bed and did not break the fuel up into sufficiently small pieces.
- The large pieces of fuel combusted on the outside only, resulting in CO release from the O_2 starved core.
- The CO combusted above the bed leading to a very high freeboard temperature, but owing to the lack of oxidation in the bed, the bed temperature was low.
- To maintain the bed temperature, the gas lances were switched on. These increased the bed temperature locally but further starved the large sludge particles of oxygen.
- The sludge feeder continued to add more fuel into the bed. This was neither mixed well nor completely combusted.
- The reducing conditions (oxygen starved atmosphere) created in the bed led to the agglomeration of ash which, in turn, caused the bed to defluidise.
- More air could not be introduced into the system as the induced draft and forced draft fans were under designed and already operating at full capacity.
- Water sprays were turned on to reduce the freeboard temperature as the design limit on the freeboard temperature is 900°C to prevent structural damage,

Refurbishment of the unit included 6 underbed nozzle feeders to distribute the feed evenly into the bed and ensure smaller feed particles. Underbed feeding is advantageous for fluidised beds owing to increased devolatilisation of the solid particles in the bed, rather than above it. When fuel particles devolatilise on the surface of the bed overbed combustion can occur, leading to the loss in bed temperature and the runaway freeboard temperatures seen above.

The feed to the incinerator is a mixture of primary and secondary sludge. The dewatering capability of primary sludge is far better the secondary sludge as secondary sludge is primarily composed of microbial cells which have a high proportion of bound water which cannot be mechanically removed. Hence primary sludge is a superior fuel. A well running WWTP should produce more primary sludge than secondary. However, if the primary settlers are not working properly, then a higher BOD load goes to the activated sludge vessels, resulting in higher biomass yields. A rough guideline to predicting the performance of the incinerator based on the relative content of primary and secondary sludge content is shown in Table 3-4.

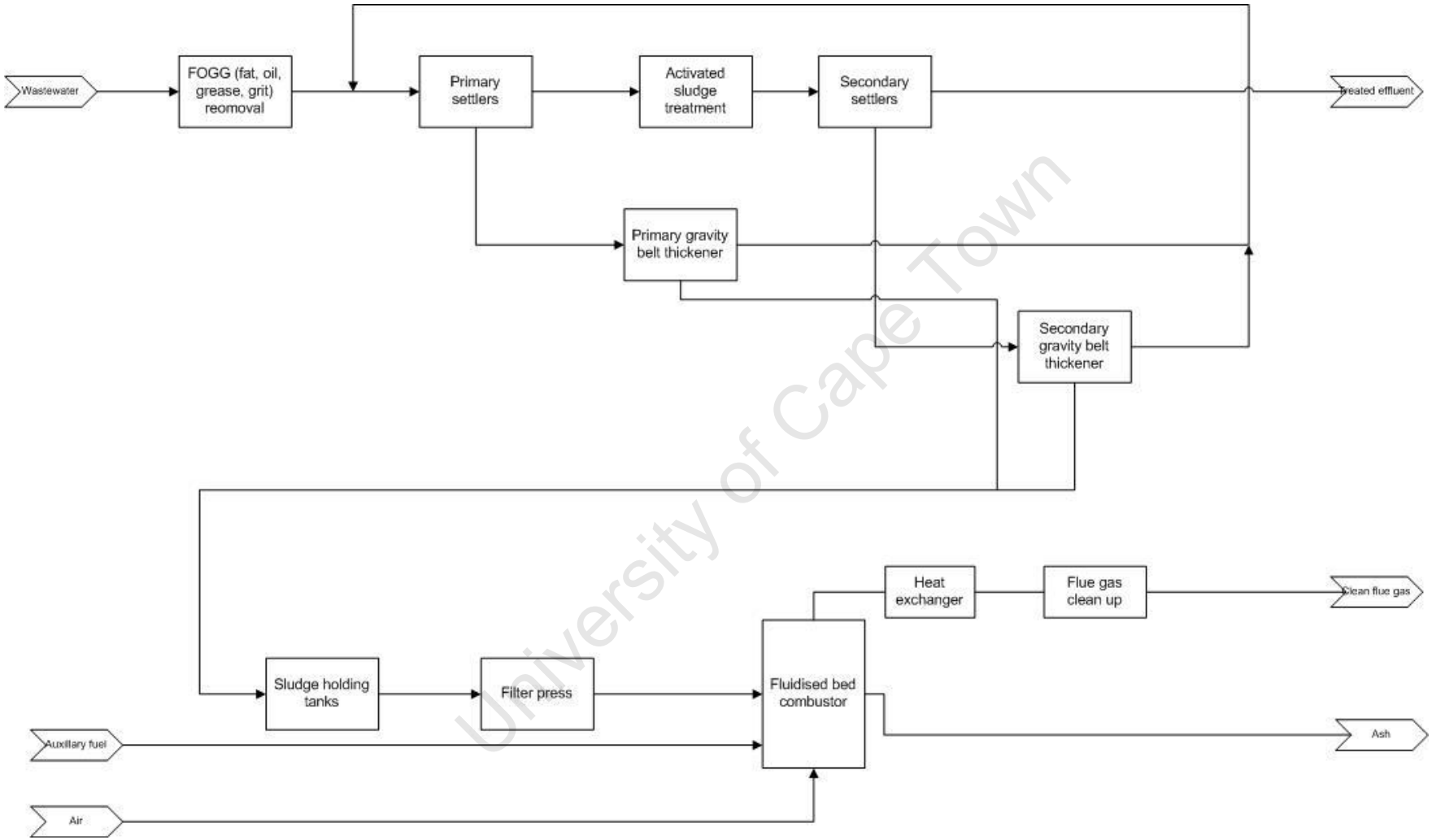


Figure 3-2: Process block diagram of Beckton WWTP

Figure 3-3: Aerial view of Beckton WWTP (obtained from maps.google.co.uk, 2008)

Table 3-4: Effect of primary: secondary sludge ratio on combustor performance (Fountain, 2007)

Primary	***Secondary***	***Operation***
60%	40%	Good
50%	50%	Acceptable
40%	60%	Problematic

3.3.2. Reading WWTP

3.3.2.1. Process description

The design of the WWTP plant at Reading, UK (commissioned in 2005) required careful consideration of visual aesthetics and odour control due to its proximity to a large shopping centre. Figure 3-4 shows the block flow diagram of the Reading WWTP. To this end, the primary settlers and gravity belt thickeners are fully covered with an extractor system drawing odorous air to a central odour control facility. The digesters are clad in light grey plastic panels to improve aesthetics. Figure 3-5 shows the egg-shaped digesters in the centre of the photograph.

The wastewater enters the plant and passes through the grit removal lanes before entering the lamella tanks, seen as the large rectangular building in the foreground. These primary settling tanks have sloping "lamellas" inside the tank to increase the surface area on which solids can settle, and channel their fall to a central sludge collection drain. The overflow of the lamellas passes to the Biological Nitrification Reactor (BNR) while the underflow passes to the primary gravity belt thickeners.

The BNR's are long channel reactors with aerobic and anoxic zones which allow nitrifying and denitrifying bacteria, respectively, to convert ammonia to nitrate and nitrate to nitrogen. In the aerobic zones, BOD reduction occurs. The flow passes into the secondary settlers at the end of the channel. The overflow from the thickeners is sent to a tertiary settler before being released to the river. About 80% of the sludge in the underflow is returned to the BNR's and 20% passes to the secondary gravity belt thickeners and the overflow is sent to tertiary settlers before being returned to the river.

The gravity belt thickeners thicken the sludge to 6% dry solids. The thickened sludge is sent to the pasteurisation unit where it is heated to 55°C for half an hour using a heat exchange surface inside the vessel. This process kills vegetative pathogens, resulting in a log 5 pathogen reduction, but does not give any improvement in the digestibility of the sludge (Shana, 2005).This may be due in part to the high content of volatile fatty acids in the primary sludge which would partially hydrolyse the sludge prior to pasteurisation.

The pasteurised sludge is then pumped to the egg-shaped digesters. The shape of these concrete vessels aid mixing, as well as ease of sludge withdrawal. The digesters are mixed by three impellers which have been found to give good radial mixing but limited axial mixing. The resulting vertical stratification of the digester contents will be minimised by the planned installation of recirculation pumps.

Biogas collects at the top of the reactor and passes to a double-walled rubber gas collection "balloon". The inner balloon collapses and expands as the gas flow fluctuates. The outer protective balloon, maintained at a pressure slightly above atmospheric, facilitates gas flow out of the inner balloon. The gas is used in a modularised Combined Heat and Power (CHP) engines which each produce 0.5 MW of electricity. The exhaust gases from the engine are used in a heat exchanger to heat the sludge in the pasteurisation unit.

The digester sludge is dewatered to ~20% solids in a centrifuge, whilst the dewatered cake is stored in silos ready to be taken away to land. The overall volatile solids destruction of the digesters is 45% compared to the 65% volatile solids reduction at Chertsey WWTP.

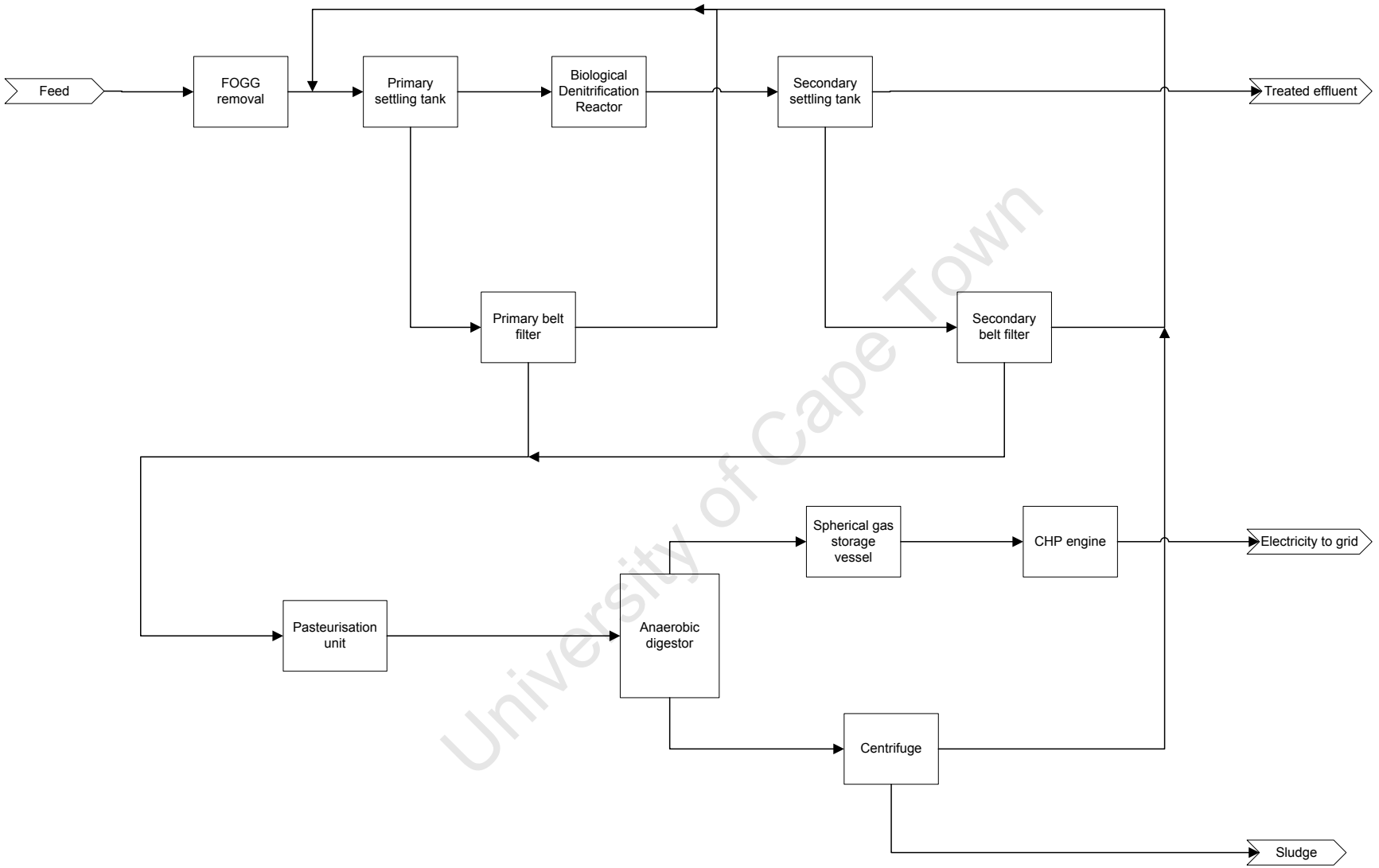


Figure 3-4: Process block diagram of Reading WWTP

3.3.2.2. Challenges to plant operation

The plant has not experienced any major operational problems. The heat exchange surfaces of the pasteurisation unit must be periodically cleaned using a concentrated H_2O_2 solution, as they become fouled with "baked" sludge. It is for this reason Thames Water operations prefer steam injection heating systems.

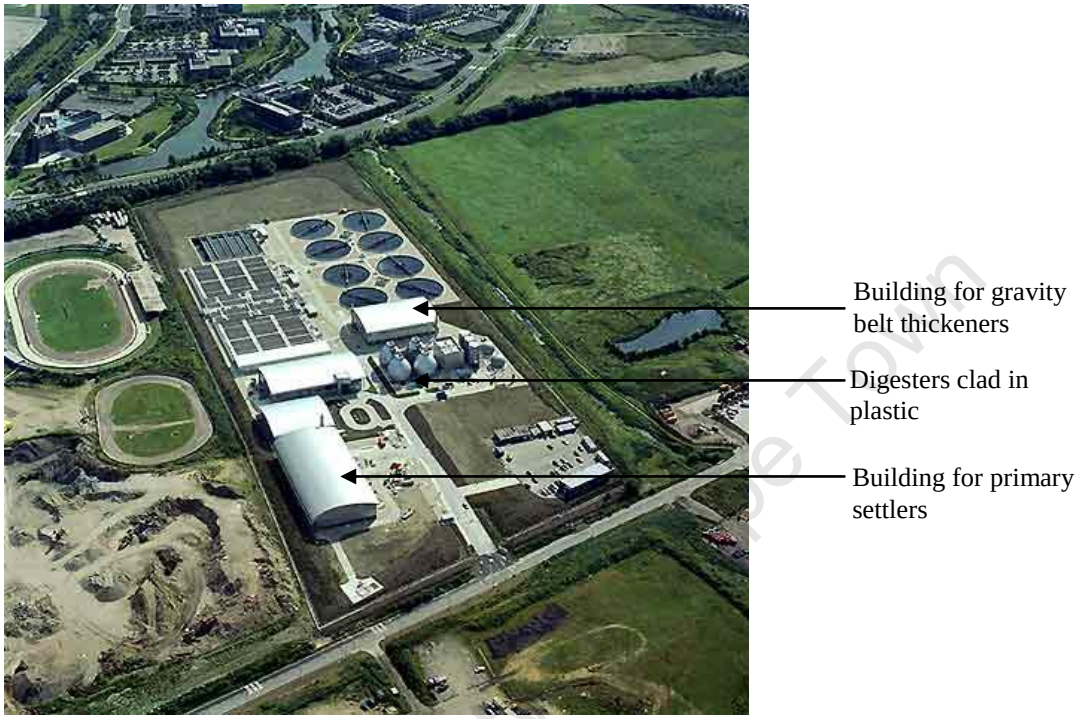


Figure 3-5: Aerial view of Reading WWTP

3.3.3. Chertsey WWTP

3.3.3.1. Process description

The Chertsey WWTP located within London, UK is smaller than Beckton and Reading. The plant features trickle bed filters and employs the CAMBI^{TM} thermal hydrolysis process to pretreat the sludge prior to anaerobic digestion. Figure 3-6 shows the block flow diagram of the Chertsey WWTP.

Wastewater enters the plant and passes through a strainer where large solids are removed and sent to landfill. Solids from storm water run-off are also separated out before the water enters the primary settlers. The primary settlers at Chertsey are circular vessels with a rotating arm to remove scum, and a base that slopes to a central sludge collection point. Sludge is then pumped to a sludge holding tank from which the overflow is passed to the trickle beds filters

and then on to the secondary thickeners. The overflow from the secondary thickeners is released to the environment and the underflow thickened in gravity belt thickeners before entering the sludge holding tank. The trickle bed filters at Chertsey WWTP offer a low energy alternative to the activated sludge process used at the Beckton and Reading WWTP's.

Chertsey WWTP accepts "import" sludge from other plants which do not have their own processing capabilities. The "import" sludge is pumped from the tanker into the sludge holding tank. The sludge from the sludge holding tank is passed through a rag remover to prevent blockage of the downstream progressive cavity pumps downstream. The sludge is dewatered to ~15% solids in a belt press. Pumping a 15% dry solids sludge is difficult and hence the dewatered sludge is re-slurried to 12% to allow for ease of pumping. The sludge is then passed to the CAMBI^{TM} thermal hydrolysis plant.

The thermal hydrolysis plant consists of three units: a pulper, a reactor and a flash vessel and operates in sequential batch mode. The sludge is pre-heated to 90°C in the pulper by injecting recycled steam from the reactor and flash tank. The sludge is then pumped to the reactor at 165°C and 8 bar where it is held under these conditions for 30 minutes, destroying the cell wall, and making the sludge easier to digest. The severe conditions also destroy pathogens and the process achieves a pathogen reduction of log 7 (Shana, 2006). The addition of the steam to the pulper and reactor reduce the solids content further to ~9%.

The pressure on the vessel is then slowly relieved, and the sludge passes to the flash vessel operated at 105°C by flashing steam back to the pulper. Any vapours from the pulper are liquefied in a dedicated offgas treatment unit and pumped, with the hydrolysed sludge, to the digester, thereby eliminating odour.

The digester is operated at a solids content of ~7% and is mixed by sparging the vessel with the biogas. The digester achieves a high volatile solids destruction of 65% as a result of the pretreatment process. The digested sludge is dewatered off site, due to capacity constraints, and the remaining sludge is sent to land.

The biogas is stored in a steel expandable gas storage vessel. Approximately 55% of the gas is used in a steam boiler to raise steam for the plant and the rest is flared. The reason for not

installing a CHP engine is that the facility to export power onto the grid is not available and installing it is a large capital expenditure with a long return on investment.

3.3.3.2. Challenges to plant operation

The thermal hydrolysis plant at Chertsey was commissioned in 1999 to increase the throughput of the existing digesters. The plant initially experienced problems with foul gas produced from the flash tank. It was re-commissioned in 2005 by CAMBI™ and the operation of the plant was also out-sourced to CAMBI™. As CAMBI™ is paid per tonne solids processed, there is a high incentive for maintaining the plant. Since re-commissioning, the plant has operated successfully with the overall result that the anaerobic digesters achieve 65% volatile solids destruction compared to the 45% seen at Reading.

Problems with the mixing of the digester at Chertsey WWTP result in acidified regions forming in the digester. Should the entire digester contents become acidified, the whole digester the unit may have to be drained and restarted; a process which could up to 3 weeks if there is no seed culture or 10 days if a seed culture is available. In the case of the Chertsey WWTP, the problem was diagnosed early, allowing remedial action to be taken *viz.* diluting the reactor contents.

This problem occurred due to an under-designed gas sparging system for sludge, which contained 7% dry solids. The pumps for the system were not able to operate all the nozzles simultaneously, so a sequential operation of nozzles was introduced such that four out of the six nozzles would operate sequentially at full capacity at any one time. This modification improved the overall mixing of the digester.

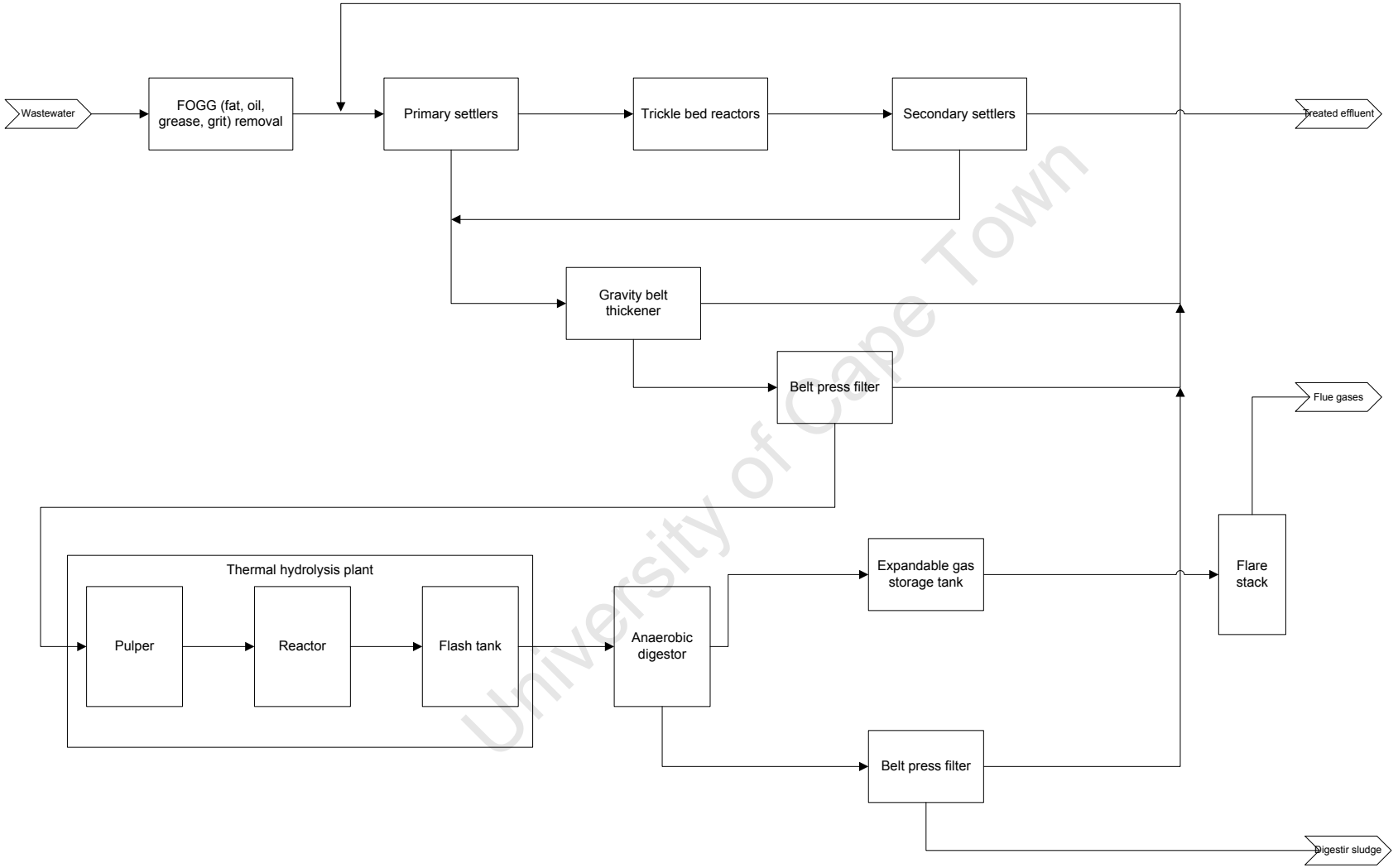


Figure 3-6: Process block diagram of Chertsey WWTP

3.4. Energy generation potential and plant-wide energy requirements

3.4.1. Energy generation

The energy generated on a dry solids basis via combustion or anaerobic digestion may be calculated from Equation 3-1 and Equation 3-2, respectively. In Equation 3-1 the energy yield from combustion is modelled as the difference between the heat released through combustion of solids (as given by the lower heating value of the fuel) and the heat required for evaporating the moisture. In Equation 3-2, the energy yield from anaerobic digestion is modelled as the product of the extent of volatile solids destruction, specific yield of methane and the lower heating value of the methane. The specific biogas yield is reported by Thames as 1 m^3/kg volatile solids at STP. The methane content of the biogas is approximately 65%, which equates to a specific methane yield of 0.65 m^3/kg volatile solids.

$$E_{combustion} = m_{DS} \cdot \left(LHV_{fuel,db} - \frac{(1-DS)}{DS} \cdot \Delta Hvap_{water} \right)$$ **Equation 3-1**

where $E_{combustion}$ = energy yield from combustion [MJ/day]
m_{DS} = mass flowrate of dry solids [kg/day]
DS = dry solids content (%)
$LHV_{fuel,db}$ = lower heating value of fuel on dry basis [MJ/kg]
$\Delta Hvap_{water}$ = heat of evaporation of water at 25°C [2.38 MJ/kg]

$$E_{AD} = m_{DS} \cdot (1-ash) \cdot VSd \cdot Y_{CH_4} \cdot LHV_{CH_4}$$ **Equation 3-2**

where E_{AD} = energy yield from anaerobic digestion [MJ/day]
m_{DS} = mass flowrate of dry solids [kg/day]
VSd = percentage volatile solids destruction (%)
ash = ash content in fuel on dry basis (%)
Y_{CH4} = yield of methane STP [0.65 m^3/kg VS]
LHV_{CH4} = lower heating value of CH_4 at STP [35.8 MJ/m^3]

3.4.2. Energy requirements

The main consumers of energy on a wastewater treatment plant are the inlet pumps and the compressors.

- The inlet pumps transport the entire input volume up from the sewer, typically at some depth, to the plant itself
- The compressors generate compressed air for the activated sludge process

Smaller energy consumers are the sludge pumps which bring the filter press or thermal hydrolysis tank to pressure. The term "parasitic load" is used to refer to these energy requirements as well as any further building energy requirements.

3.4.2.1. Pumps

The energy requirement for pumping the wastewater up from the sewer into the plant is shown in Table 3-5. Here it is estimated that 3.4 kW per ton solids is required to deliver wastewater with a solids content of 0.03% from a 7 m deep sewer.

At Beckton WWTP, the filter press was pressurised to 6 bar by progressive cavity pumps and up to 10 bar, piston pumps were used. The piston pumps at Beckton WWTP require more maintenance than the progressive cavity pumps. At Chertsey WWTP progressive cavity pumps are used throughout the plant. The energy requirement for pressurising the sludge to 10 bar in the filter press is estimated as 0.24 kW per ton dry solids, as shown in Table 3-6. Progressive cavity pumps at Chertsey pump a 12% dry solids content sludge to a pressure of 8 bar in the thermal hydrolysis tank. The energy requirement for this process is 0.10 kW/ton dry solids as shown in Table 3-6.

Table 3-5: Energy requirement to pump wastewater up from sewer

Solids flow into plant	ton/day	1
Solids content of flow [a]	% w/v	0.03%
Wastewater flow into plant	m^3/day	3,333
Sewer depth [a]	m	7
Static head	Pa	70,000
Shaft energy requirement	kW	2.7
Pump efficiency	%	80%
Pump energy requirement	kW	3.4

a.) Thames Water practice (Fountain, 2008)

Table 3-6: Energy requirement to pressurise sludge

Energy requirement for:		Filter press	Thermal hydrolysis tank
Solids flow to filter press	ton/day	1	1
Solids content prior dewatering	% w/v	6% [a]	12%
Flow to pumps	m^3/day	17	8
Pressure increase	bar	10	8
Shaft energy requirement	kW	0.19	0.08
Pump efficiency	%	80%	80%
Pump energy requirement	kW	0.24	0.10

a.) Thames Water practice (Fountain, 2008)

3.4.2.2. Compressors

In order to realise the gas-liquid oxygen transfer required for the activated sludge process, air under pressure is sparged into the tanks. Energy requirements for aerating a tank range from 0.5 to 1 kWhr/m^3 (Ullmans, 2003a). In Table 3-7, a specific aeration energy requirement of 5 kW/ton dry solids is calculated according to the activated sludge process flow scheme shown in Figure 3-7. The trickle filter process requires no aeration and therefore no energy.

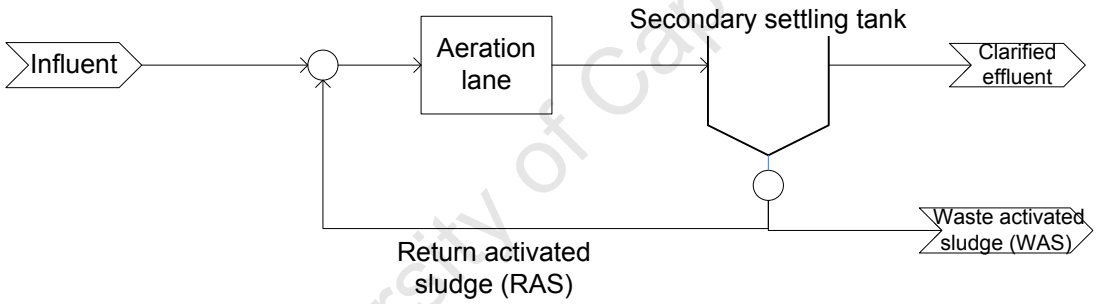


Figure 3-7: Activated sludge process on which aeration energy calculation based

Table 3-7: Aeration energy requirements for the activated sludge process

Solids flow to aeration lanes	ton/day	1
Influent dry solids content [a]	% w/v	0.8%
Concentration of thickened sludge [a]	% w/v	4%
WAS:RAS [b]	1: 4	4
Influent flow	m^3/day	125
WAS flow	m^3/day	25
Flow through aeration lanes	m^3/day	150
Specific aeration energy requirement [c]	kJ/m^3	2,700
Aeration energy requirement	kW	5

a.) Waites *et al.*, 2001
b.) Thames Water best practise (Fountain, 2008)
c.) Ullmans, 2003a

3.5. Comparison of technologies

3.5.1. System definitions and system boundaries

The basis for the calculations was chosen as 1 ton dry solids/day. The scenarios chosen for analysis are combustion (or incineration) of 20% dry solids sludge; anaerobic digestion of 10% dry solids sludge and anaerobic digestion with thermal hydrolysis pretreatment of a 10% dry solids sludge. Associated with these are different volumetric flow rates.

The system boundaries for the energy balances were chosen such that internal heat recovery loops (if present) could be ignored. Sensible heat losses arise when releasing a warmed stream to the environment.

- Combustion: The air stream enters at 15°C and is pre-heated by the steam turbine exhaust and warm flue gases. The flue gases are released at an assumed temperature of 105°C. The lower heating value of a fuel is given at a reference temperature of 25°C and hence sensible heating requirements of the sludge do not need to be factored in where the inlet temperature of the sludge is approximately 25°C. The system boundary is shown in Figure 3-8.
- Anaerobic digestion: The sludge stream enters at 15°C and is heated to 50°C prior to entering the digester. Further heat inputs to the digester are not necessary and the unit operates at a temperature of 37°C. The system boundary is shown in Figure 3-9.
- Enhanced anaerobic digestion: The sludge enters the thermal hydrolysis plant at 15°C and leaves at 105°C. Internal heat recovery loops can be ignored. There is no heat recovery on 105°C stream and it is cooled by cooling water to 50°C before being pumped to the digester. Again, the hot inlet stream to the digester provides heating to the entire unit. The system boundary is shown in Figure 3-10.

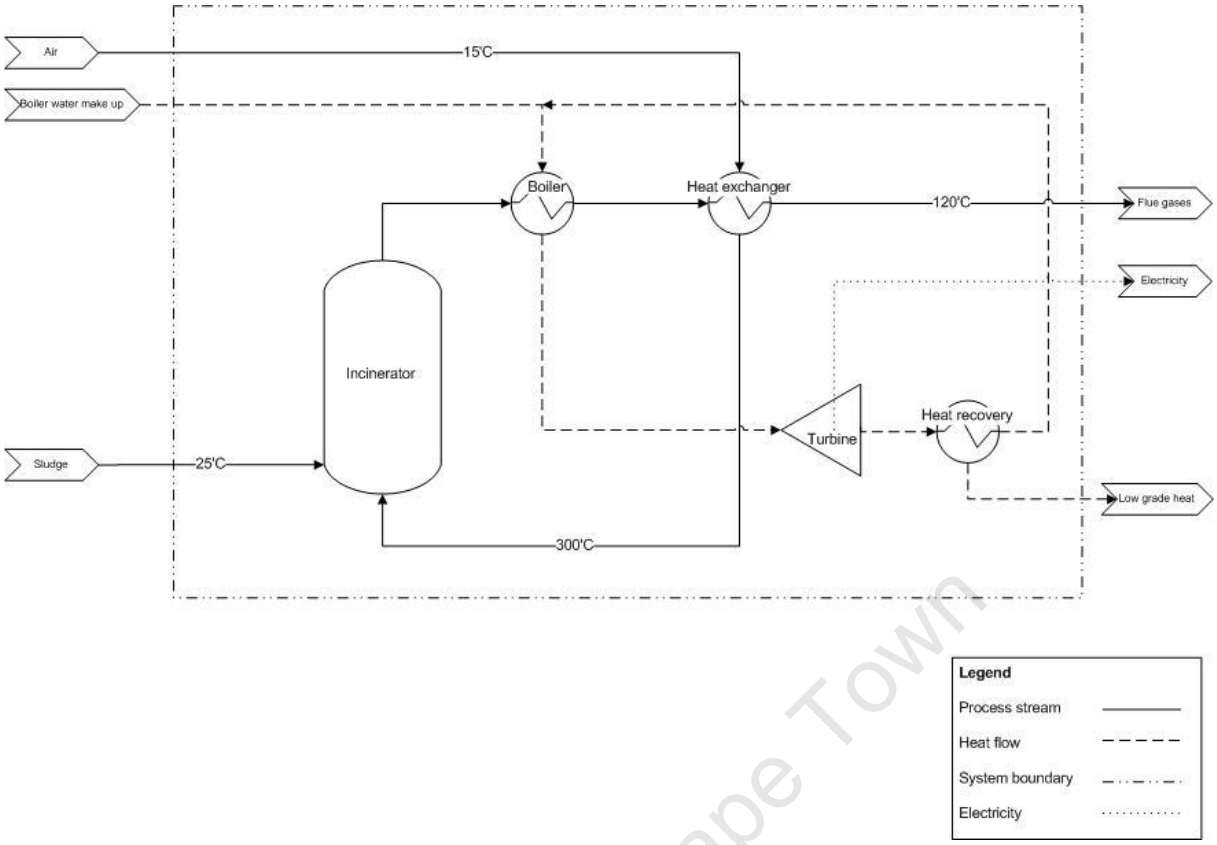


Figure 3-8: Combustion system boundaries

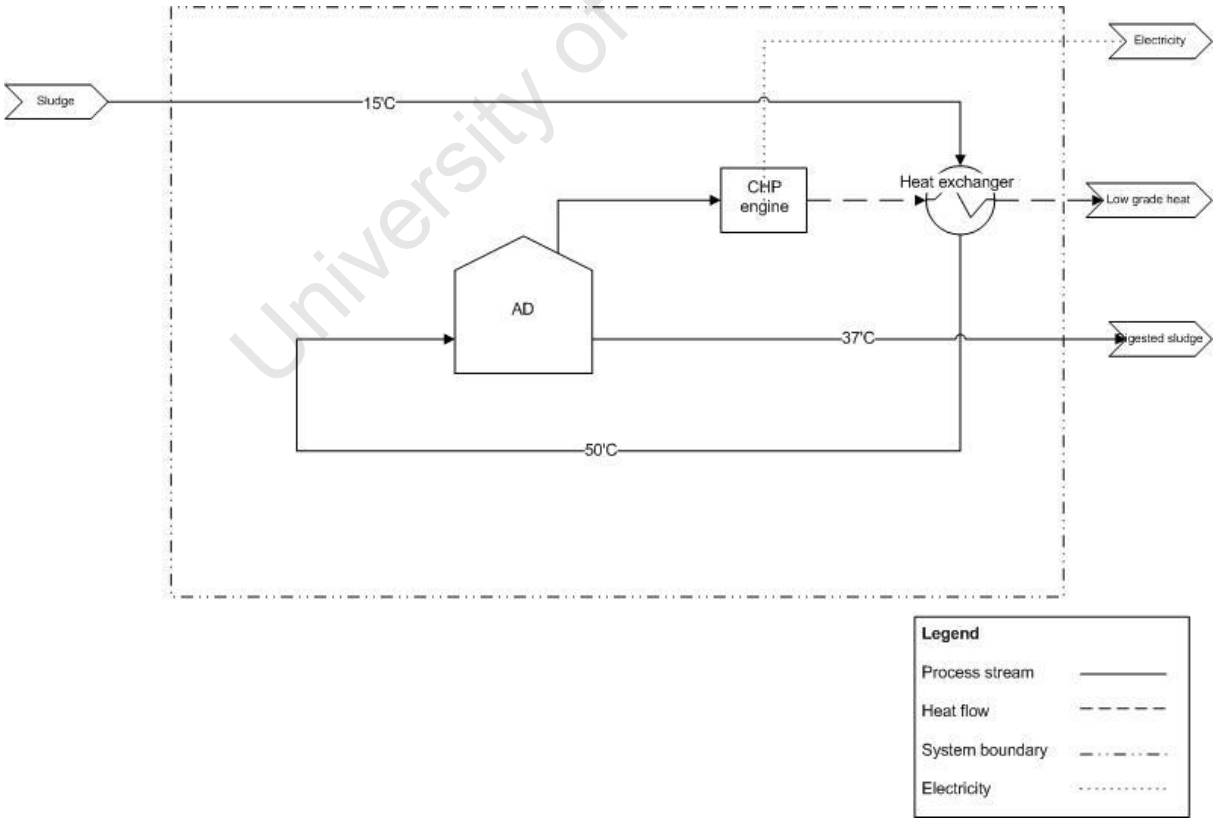


Figure 3-9: Anaerobic digestion system boundary

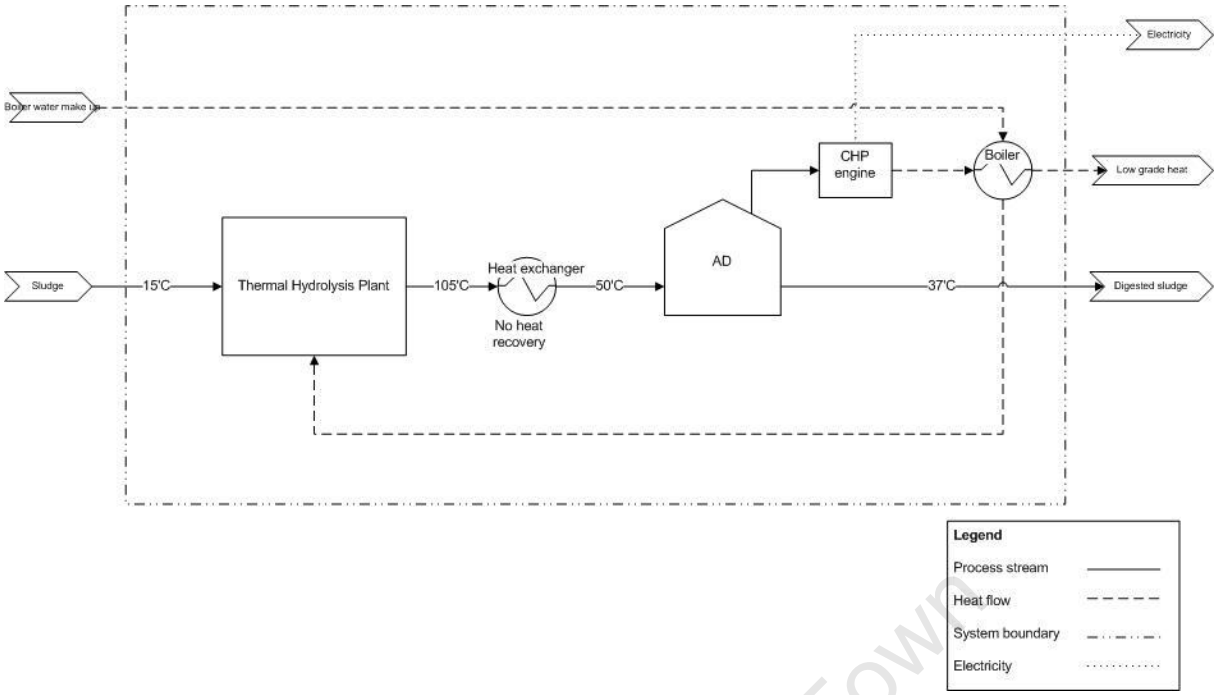


Figure 3-10: Enhanced anaerobic digestion system boundary

3.5.2. Calculations

The system definitions presented in Section 3.5.1 are used to compare the energy recovery across the three Thames Water WWTP's. The calculation is presented in Table 3-8. The gross energy production from incineration and anaerobic digestion of sludge is calculated according to Equation 3-1 and Equation 3-2, respectively. Thermal efficiencies based on Thames Water operational experience are used to calculate the high grade heat and electricity output. The difference in the efficiencies is due partly to CHP engines having higher efficiencies and partly due to the state of the art equipment used at Reading.

Table **3-9** shows the plant data for the gross electrical output and parasitic load. The high grade heating requirements, as shown in Table 3-11, are calculated for each process based on the specified system definitions. The parasitic loads for each plant, as calculated in Section 3.4.1, is shown in Table 3-10

3.5.3. Results and discussion

Table 3-8 andTable 3-9 show reasonable agreement between the calculated and actual plant data for gross electricity output exsists. Both the calculation and the plant data show that enhanced anaerobic digestion gives the highest energy yield.

Table 3-8: Calculated gross electricity and high grade heat production

Plant		**Beckton**	**Reading**	**Chertsey**
Energy recovery process		***Combustion***	***AD***	***Enhanced AD***
Solids mass flow	ton dry solids/day	1	1	1
Solids content	% w/w	20%	10%	10%
Water mass flow	ton/day	4	9	9
Ash content (dry basis) [1.]	%	33%	33%	33%
VS destruction	%		45%	65%
Methane yield/VS destruction [2.]	m^3/kg		0.65	0.65
LHV methane [2.]	MJ/m^3		35.8	35.8
LHV fuel (dry basis) [3.]	MJ/kg	19	-	-
$\Delta Hvap_{water}$	MJ/kg	2.38		
Methane production	m^3/day		196.0	283.1
Gross thermal power out	MJ/day	9,480	7,016	10,134
Gross thermal power out	kW	**110**	**81**	**117**
Residues	Amount	0.33	3.5	2.8
	Quantity	ton ash per day	ton sludge at 20% dry solids content per day	ton sludge at 20% dry solids content per day
Energy conversion unit		Steam turbine	CHP engine	CHP engine
Conversion to high grade heat	%	30%	40%	40%
	kW	**33**	**32**	**47**
Conversion to electrical power	%	30%	40%	40%
	kW	**33**	**32**	**47**

Notes

1.) This value was taken from the average dry sludge composition from Phyllis database (ECN, 2008)

2.) $LHV_{methane}$ at STP taken from Perry (1999)

3.) LHV_{fuel} taken as sewerage sludge average LHV reported by Thames Water.

Table 3-9: Gross electricty production and parasitic loads from plant data

Plant data (2007)		***Beckton***	***Reading***	***Chertsey***
Energy conversion process		Combustion	AD	Enhanced AD
Gross electricity produced/ton dry solids	kWhr/tds	619	737	1057
Parasitic load	kWhr/tds	393	366	234
Gross electric power produced	**kW**	**26**	**31**	**44**
Parasitic load	**kW**	**16**	**15**	**10**

Table 3-10: Parasitic loads

Pump up from sewer	kW	3.4	3.4	3.4
High pressure pumps	kW	0.24		0.10
Compressors	kW	5	5	0
Total	kW	**8.3**	**8.1**	**3.5**

Table 3-11: Calculation of high grade heating requirements

		Air to combustor	*Sludge to pasteurisation*	*Sludge to reactor*
Carbon content of sludge (dry basis) [1.]	mass %	30%		
Stoichiometric air requirement	kmol/day	119		
% O_2 in flue gas [2.]	%	6%		
Excess air	%	40%		
Air flow to combustor	kmol/day	167		
Cp_{air} [3.]	kJ/kmol.K	31		
Cp_{water} [4.]	kJ/kg.K		4.2	4.2
T_{in}	°C	25	15	15
T_{out}	°C	120	50	105
ΔT	K	95	35	90
$Q_{sensible}$	kW	**6**	**17**	**44**

Notes
1.) Taken as average C content from Phyllis database
2.) Design parameter
3.) Average Cp for temperature range 300°C to 750°C
4.) Assume $Cp_{sludge} = Cp_{water}$

From Table 3-8 and Table 3-11 it can be seen that is all cases the high grade heat generated in the process exceeds the high grade heat requirements. In the case of combustion and anaerobic digestion, the high grade heat output exceeds the requirements by five times and double, respectively. This brings into question whether the high grade heat could better be used on these plants, e.g. sludge drying prior to incineration or pasteurisation at higher temperature to effect greater sludge hydrolysis. The plants' process heat and electrical energy inputs and outputs are summarised in Figure 3-11. The calculated parasitic loads are ~7 kW lower than that seen on the plant, a difference which could be attributed to electricity requirements other than process requirements, *e.g.* office lighting and heating.

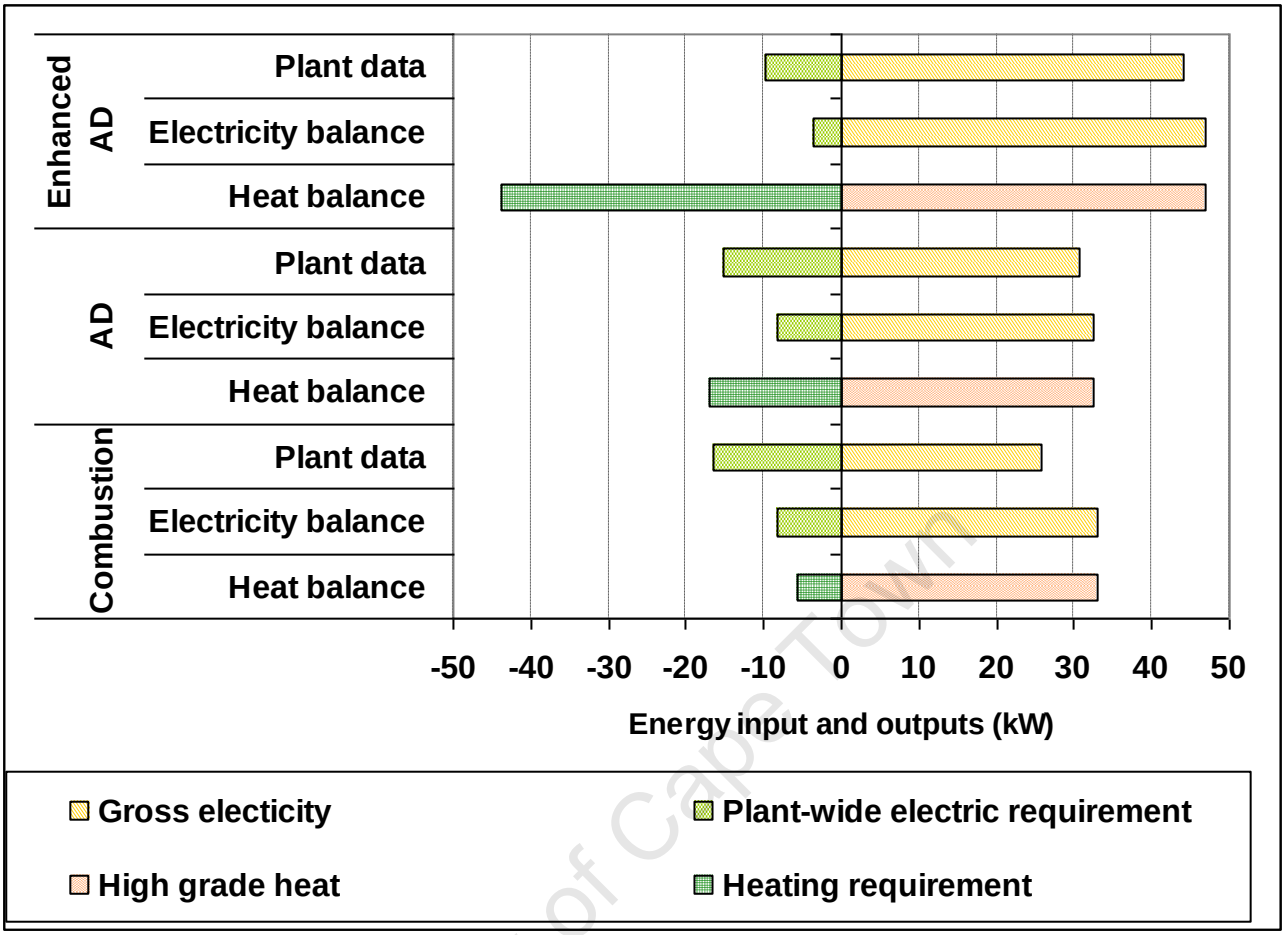


Figure 3-11: Summary of calculated and actual electricity balances and heat balance for the three processes

To investigate the lack of agreement between calculated values and plant data for combustion, the sensitivity of the calculation to dry solids content was investigated. This was chosen owing to the variability in the dry solids content of the filter cake. The dependence of gross thermal energy produced on dry solids content is shown in Figure 3-12. Energy production based on two heating values is presented: 19 MJ/kg, the value reported by Thames Water and 13 MJ/kg, an average value of data from Phyllis database (ECN, 2008). The Thames Water value is atypically high when compared to the range of values in the Phyllis database. From Figure 3-12, it can be seen that for a sludge with an LHV of 13 MJ/kg, the gross thermal energy output drops from 75 kW/ton dry solids to 0 kW/ton dry solids as the solids content decreases from 30 to 15%. For a sludge with an LHV of 19 MJ/kg, the gross thermal output drops from 155 kW/ton dry solids to 0 kW/ton dry solids as the solids content decreases from 30% to 12%.

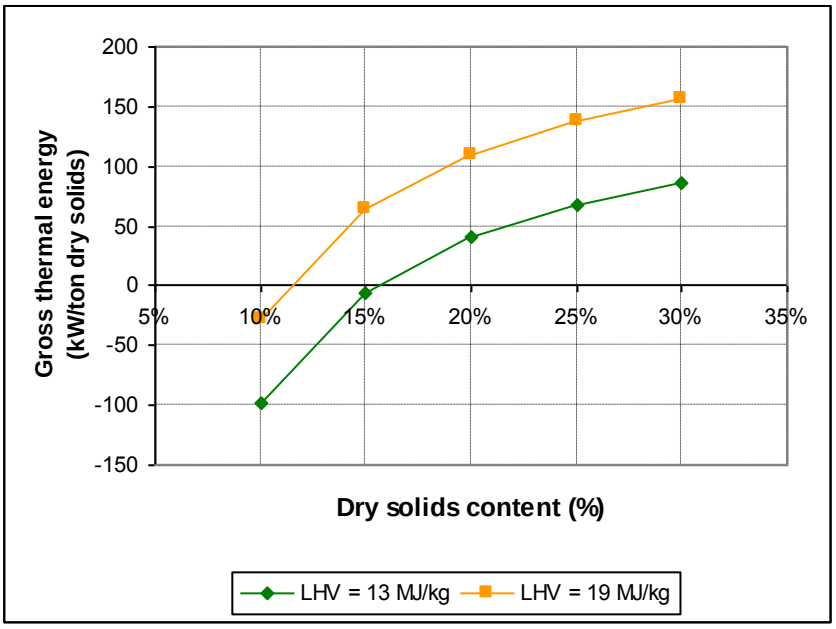


Figure 3-12: Sensitivity of gross thermal energy production by combustion on solids content

Whilst greatest energy recovery is achieved using enhanced anaerobic digestion, a disadvantage is larger residue production. Anaerobic digestion results in a residue of 3 to 4 tons of digested sludge as opposed to 0.3 tons of ash. This has to be dealt with by further dewatering and then sending it to land. Enhanced anaerobic digestion produces less digested sludge which has cost advantages for final disposal to land. A further advantage of enhanced anaerobic digestion is the improved dewaterability. As the cells have been burst open, the intracellular water is released and can be removed by mechanical dewatering techniques.

Not only is enhanced anaerobic digestion superior from an energy recovery point of view, it is also superior economically in terms of economy of final sludge disposal. Preliminary estimates of the nett process costs are shown in Table 3-12. The calculation takes into the operating cost of the process, the cost of final disposal and income from sale of electricity to the grid.

Table 3-12: Relative costs of different sludge disposal options (Fountain, 2008)

Process option	***Cost (£/ton sludge)***
Liming and putting to land	£60
Incineration	£30
Anaerobic digestion	£0
Enhanced anaerobic digestion	-£20

3.6. Conclusions

The analysis carried out on the Thames Water WWTP's shows that enhanced anaerobic digestion, in this case with pretreatment by thermal hydrolysis, is the best option in terms of energy recovery and economics. The heat balance for enhanced anaerobic digestion showed that heat requirement is met by the high grade heat generated in the CHP engine. The heat generated in anaerobic digestion without pretreatment and combustion exceeded the heating requirements indicating potential for alternate uses of waste heat, e.g. sludge drying.

The equations used to derive the energy recovery gave results which compared well to plant data for anaerobic digestion. The deviation from plant data in the case of combustion was attributed to the non-linear dependence of dry solids content on energy yield, *i.e.* a small decrease in dry solids content results in a large drop in energy yield. A sensitivity analysis on the factors affecting anaerobic digestion showed a linear dependence of gross thermal energy output on volatile solids destruction and ash content.

Chapter 4 Biogenic waste feedstock characteristics and their effect on energy recovery

In this Chapter, two studies are carried out in which the effects of biogenic feedstock characteristics on energy recovery are considered. Section 4.1 considers the effect of biochemical feedstock characteristics on energy recovery in an anaerobic digester. The energy costs of anaerobic digestion pretreatments are also considered. Section 4.2 considers the effect of volatile combustion in bubbling fluidised beds.

4.1. Effect of recalcitrant substances on biogas yield

4.1.1. Biochemical components affecting digestibility

4.1.1.1. Lignocellulosic components

Cellulose, hemicellulose and lignin form the structural components of higher plants. The biodegradability of agricultural residues, forestry residues and bagasse are affected by these components. A short review of the structure of cellulose, hemicellulose and lignin and their enzymatic hydrolysis is presented.

(1) Cellulose

Cellulose consists of up to 10 000 D-glucose molecules linked by β,1-4 glycosidic bonds (Bailey and Ollis, 1986). Cellulose molecules align to form microfibrils which consist largely of crystalline cellulose regions broken occasionally by amorphous cellulose regions. The crystalline cellulose creates an extensive level surface which is difficult for hydrolytic enzymes to attach to (Schwarz, 2001). Hydrogen bonding between and within the layers strengthens the microfibril (Schwarz, 2001; Bailey and Ollis, 1986). The microfibrils usually exist in a matrix of other polymers such as hemicellulose and lignin which further prevent hydrolytic enzyme binding.

Cellulose degradation is important in terms of feasibly processing lignocellulosic feedstocks. A number of components must work together to hydrolyse cellulose, as illustrated in Figure 4-1.

- A non-catalytic carbohydrate binding module (CBM) to attach to the substrate and position the catalytic component
- Endoglucanases to degrade amorphous cellulose regions and to cut into long cellulose strands
- Exoglucanases or cellobiohydrolases progressively cut from both ends of shorter cellulose strands
- Cellobiases and cellodextrinases are specific β-glucosidases and degrade the resulting oligosaccharides or disaccharides

The cellulase system of the aerobic fungi, *Trichoderma reesei*, consists of a single cellulase connected to a CBM. The cellulases compete for a limited number of binding sites on the cellulose surface. The competition is marked by a reduction in synergy between enzymes if one component is increased. Thus effective hydrolysis is dependent on both the amount of enzyme and the correct ratio of components (Schwarz, 2001).

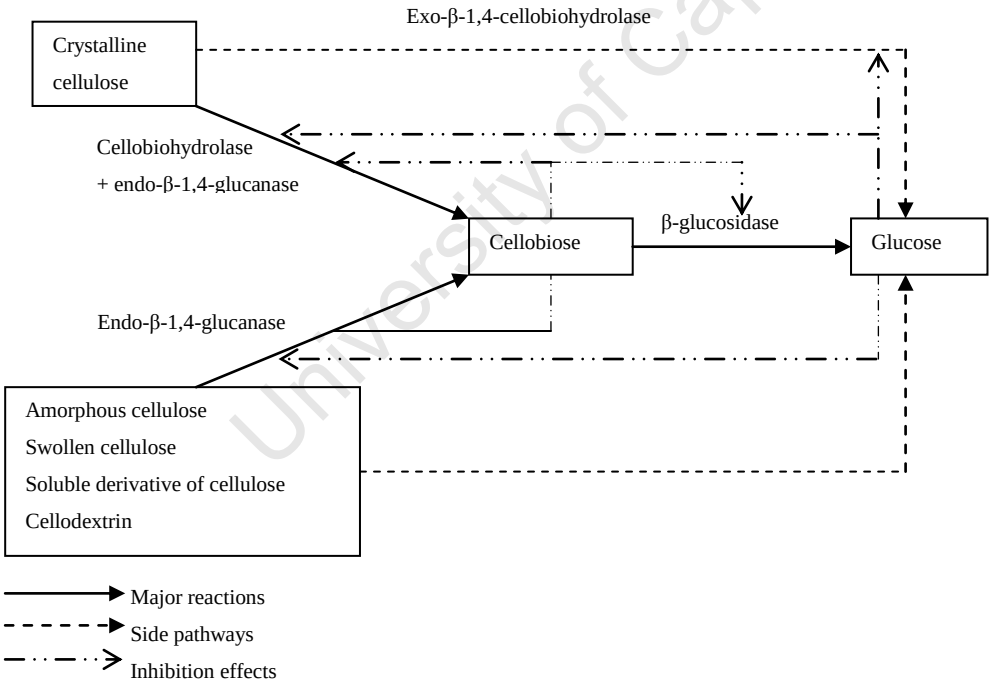


Figure 4-1: Cellulase system of *Trichoderma reesei* (Bailey and Ollis, 1986)

In contrast, anaerobic rumen bacteria such as *Clostridia* and *Ruminococci* exhibit a multi-enzyme complex termed a cellulosome. The cellulosome is a cell protuberance and consists of a CBM with lower specificity and the correct ratio of enzymes for effective hydrolysis. This system has a number of advantages (Schwarz, 2001):

- Synergy between the components is optimised by having the correct ratio
- Non-productive binding is avoided by optimal spacing of the components
- Competition for limited binding sites is reduced as all the components are bound through a single domain with low specificity
- A halt in hydrolysis on depletion of one type of cellulose is prevented by the presence of other enzymes with different specificity.

(2) Hemicellulose

Hemicellulose consists of short branched polymers of pentoses (xyloses and arabinose) and hexoses (glucose, galactose and mannose) (Bailey and Ollis, 1986). Most hemicelluloses are β-1,4 linked except for galactose containing hemicelluloses which are β-1,3 linked. Unlike cellulose, hemicellulose is not chemically homogenous. Hardwood hemicelluloses contain mostly xylans, whereas softwood hemicelluloses contain mostly glucomannans (Saha, 2003). Hemicellulose is hydrogen bonded to cellulose providing structural support to the plant and protecting the cellulose from enzyme attack (Mosier *et al.*, 2005).

The heterogeneous nature of hemicellulose means that many enzymes are involved in its hydrolysis. Details of the enzymes involved in the hydrolysis of heteroarabinoxylans are given in Table 4-1. Whilst many enzymes with different specificities are involved, the enzymatic hydrolysis of hemicellulose is not as difficult tightly packed crystalline structures are not present (Saha, 2003). Alternatively, an acid pretreatment may be used to hydrolyse the hemicellulose to its sugars. Fermentative or acidogenic bacteria can convert the mixed sugars arising from hemicellulose hydrolysis to organic acids, including acetate, propionate, succinate and lactate (Saha, 2003).

Table 4-1: Enzymatic hydrolysis of heteroarabinoxylans (Saha, 2003)

Enzyme	***Mode of action***
Endo-xylanase	Hydrolyses mainly interior β-1,4-xylose linkages of the xylan backbone
Exo-xylanase	Hydrolyses β-1,4-xylose linkages releasing xylobiose
β - Xylosidase	Releases xylose from xylobiose and short chain xylo-oligosaccharides
α - Arabinofuranosidase	Hydrolyses terminal non-reducing α-arabinofuranose from arabonoxylans
α - Glucuronidase	Releases glucuronic acid from glucuronoxylans
Acetylxylan	Hydrolyses acetylester bonds in acetyl xylans
Ferulic acid esterase	Hydrolyses feruloylester bonds in xylans
ρ - Coumaric acid esterase	Hydrolyses ρ-coumaryl ester bonds in xylans

(3) Lignin

Lignin is a polyphenolic compound with no set repeated substructure. The three monomers which make up lignin are p-coumaryl alcohol, coniferyl alcohol and sinapyl alcohol. Coniferyl alcohol is found predominately in softwoods (*e.g.* pine, fir, spruce) and both coniferyl and sinapyl alcohols are found in hardwoods (or broad-leaved trees). p-Coumaryl alcohol is a minor component of grass lignin (Helm, 2000). The lignin molecule is synthesized by a free radical coupling reaction initiated by the enzyme phenoloxidase (Crawford, 1981). There are three most probable sites for free radical coupling between the monomers, allowing for a highly irregular structure to be synthesised. This irregularity yields it very resistant to enzyme hydrolysis.

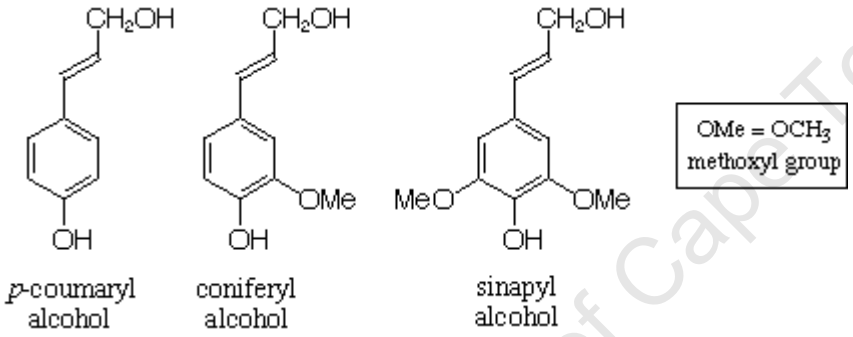


Figure 4-2: Lignin monomers (Helm, 2000)

The most common bond, present in 40 to 60% of all inter-unit linkages is the β-O-4 ether bond (Helm, 2000). This is the bond which is usually broken during delignification processes in the pulp and paper industry and the cellulosic ethanol industry. Delignification generally results in 30 to 50% solubilisation of biomass but is necessary to expose the cellulose (Crawford, 1981). There exists a number of lignin degrading micro-organisms *e.g.* white-rot fungi, brown-rot fungi, soft-rot fungi and some bacteria (*e.g. Pseudomonas sp*), however, the process is very slow. In the case of bacterial degradation, it can take up to 15 days for a 10% mass reduction in lignin (Crawford, 1981). Thus, lignin is regarded as non-biodegradable in current enzymatic hydrolysis systems (McKendry, 2002a).

4.1.1.2. Microbial cells

Microbial sludges such as domestic sewerage sludge and algal sludge resist biological action due to the complex cellular walls of the micro-organisms which make up the sludge (Speece, 1996). Mechanical and thermal pretreatments break down the cell walls into smaller components which are more susceptible to enzyme hydrolysis. Mechanical and thermal

pretreatments also allow the intracellular proteins and carbohydrate storage products to be released into the bulk liquid to be utilised by the anaerobic digester microbes. Here, the structure of bacterial cell walls and algal cell walls are considered.

(1) Bacterial cell walls

Bacterial cell walls can be classified according to their structure by Gram staining. The bacteria retaining the purple iodine Gentian-violet complex even when washed with polar solvents are termed Gram positive and those not retaining the complex, Gram negative. Common to both types of cell envelopes is the inner peptidoglycan layer and cytoplasmic membrane. The peptidoglycan layer consists of alternating β-1,4 linked disaccharides of N-acetyl-D-glucosamine (NAG) and N-acetyl-muranic acid (NAM) (Bailey and Ollis, 1986; Middelberg, 1995; Harrison, 1991). NAG comprises a tetrapeptide side chain onto which a pentapeptide crosslink is attached, allowing crosslinking of the linear saccharide polymers. The cross-linking provides shape and tensile strength to the cell. The cytoplasmic membrane consists of a phospholipid bilayer with embedded protein molecules. The proteins regulate the movement of nutrients through the cell membrane. The cytoplasmic membrane is the biochemical boundary of the cell, controlling transfer of molecules but providing little structural strength (Harrison, 1991).

Gram positive cell walls consist of a thick peptidoglycan layer of 15 to 80 nm, compared to 2 nm for Gram negative cells (Bailey and Ollis, 1986). This surrounds the cytoplasmic cell membrane. Gram negative cell envelopes consist of 4 layers as shown in Figure 4-3. The outermost layer is a lipid bilayer comprising of transmembrane proteins, phospholipids and lipopolysaccharides (LPS) and is anchored to the peptidoglycan layer. The lipid bilayer of a Gram negative cell protects the inner layers from direct chemical attack and the peptidoglycan layer provides mechanical strength (Harrison, 1991). Although Gram positive cells do not have a lipid bilayer, their peptidoglycan layer is much thicker and hence yields them less amenable to mechanical cell disruption techniques, such a high pressure homogenisation or bead milling. Between the peptidoglygan layer and the cytoplasmic membrane is the periplasmic space which can contribute up 20 to 40% of the cell mass (Bailey and Ollis, 1986).

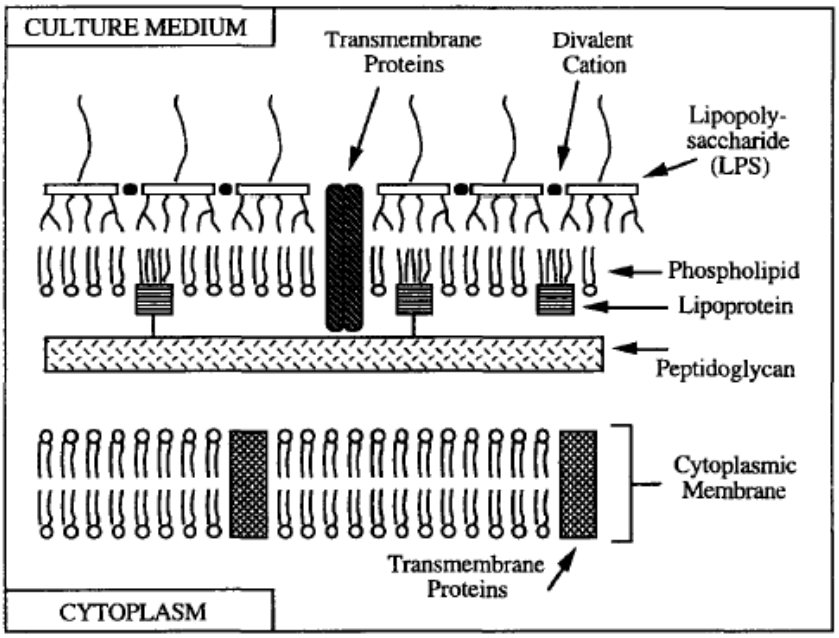


Figure 4-3: Gram-negative cell envelope (Middelberg, 1995)

(2) Algal cell walls

Algal lipids hold much promise as a feedstock for the production of second generation biodiesel. Such a facility will produce an algal biomass waste stream. The cell wall structures of three lipid producing algae genus are described; namely, Cyanophyta (under which *Dunaliella* and *Spirulina* fall), diatoms, and Chlorophyta (under which *Chlorella* and *Chlamydomonas* fall). Cyanophyta are prokaryotic, meaning they do not have membrane bound organelles (Vymazal, 1995). Their cell wall is similar to Gram negative bacteria. The cell wall of diatoms and Chlorophyta consist of a fibrillar component of, most commonly, cellulose I. Cellulose I is not as crystalline as the cellulose found in higher plants and hence is more amenable to hydrolysis. The fibrillar component is embedded in a mucilaginous matrix. In Chlorophyta the matrix consists of amorphous polysaccharides (Vymazal, 1995). The carbohydrate storage products which would be released on cell disruption are given for the different algae genus in Table 4-2.

Table 4-2: Algal storage products (Barsanti and Gaultieri, 2006)

Algal genus	***Carbohydrate storage products***
Cyanophyta	Cyanophycin starch (argine and asparagines) Cyanophycean starch
Heterokontophyta and Haptophyta (diatoms)	β-1,3- glucan
Dinophyta (diatoms)	α-1,3- glucan
Chlorophyta	α-1,4- glucan

4.1.2. Increasing energy yield from anaerobic digestion

The energy yield of anaerobic digestion is dependent on the volatile solids (organic solids) destruction and the ash content of the fuel as seen in Equation 3-2. Anaerobic digestion at a methane yield of 0.65 m^3/kg volatile solids, 100% volatile solids destruction and no ash gives a maximum energy yield of 23 MJ/kg dry solids.

4.1.2.1. Predicting volatile solids destruction

The volatile solids destruction for sewerage sludge anaerobic digesters ranges from 15 to 45% depending on the operation. In Chapter 3, it is shown that pretreatment allows greater volatile solids destruction (up to 65% with thermal hydrolysis). The volatile solids destruction of simple carbohydrate loaded streams such as those from breweries or fruit processing operations reaches 95% and higher. Lignin is not considered digestible and hence the maximum digestibility of lignocellulosic material will only be the non-lignocellulolytic components. The Biochemical Methane Potential assay was developed to determine the digestibility of a feedstock experimentally (Speece, 1996; Owen et al, 1979). A small batch anaerobic digester is inoculated with acclimatised sludge and the cumulative gas production is monitored. Initial and final COD readings are measured to give the extent of volatile solids destruction.

4.1.2.2. Energy requirements of pretreatments

The improvement in volatile solids destruction due to the action of a particular pretreatment should not be offset by the energy requirements of the pretreatment itself. In Section 0 it was seen that the energy requirement of the CAMBI^{TM} thermal hydrolysis plant at Chertsey WWTP was less than the high grade heat produced by the CHP engine did not reduce the gross energy yield; *i.e.* the improvement in volatile solids destruction due to the pretreatment was worthwhile. The effects of operational parameters on the energy requirement of two types of pretreatments are considered, *viz.* high temperature, high pressure (HTHP) treatments and high pressure homogenisation. These processes have been described in Chapter 2. The operating conditions for the processes re-iterated in Table 4-3.

Table 4-3: Operating conditions of selected pretreatments

	Pretreatment technology	Chemicals used	Batch or continuous process	Temperature (°C)	Pressure (atm)	Reaction or residence time (min)	Solids concentration (wt %)
High temperature, high pressure (HTHP) pretreatments	Dilute sulphuric acid (co current)	0.5–3.0% sulphuric acid	Continuous	130–200	3–15	2–30	10–40
	Dilute sulphuric acid (percolation)	0.07–0.1% sulphuric acid	Continuous	190–200	20–24	12–24	2–4
	Hot water hydrolysis	Water	Semi-batch	160–190	6–14	10–30	5–30
	ARP	10–15 wt.% ammonia	Continuous	150–170	9–17	10–20	15–30
	SAA	Dilute aqueous ammonia	Batch	25–60	1		
	Lime	0.05–0.15 g Ca(OH)2/g biomass	Batch	70–130	1–6	1–6 h	5–20
	Hot water hydrolysis	None	Semi-batch	165	8	30	8-12
	High pressure homogenisation	None	Continuous	Not applicable	Up to 120 MPa	Not applicable	Maximum 10%

(1) Energy requirements of HTHP pretreatments

The energy requirement for HTHP pretreatments may be calculated from Equation 4-1. The following assumptions were made in this calculation:

- The heat capacity of the feedstock slurry was equivalent to water (4.2 kJ/kg/K)
- The incoming stream is at a temperature of 15°C
- Adiabatic conditions are assumed for the duration of the holding time
- No heat recovery from the hydrolysate stream occurs

Energy requirements per ton dry solids are presented as a function of dry solids content and pretreatment temperature in Figure 4-4. Dry solids content is considered up to 40%. It is noted that for the pretreatments developed for microbial sludges, *viz.* hot water hydrolysis and high pressure homogenisation, pumping of slurries greater than 10% is difficult. The operating temperatures considered are 55°C; 120°C and 200°C. The energy requirement for pressurising the feedstock has been seen from Section 3.4.1 to be smaller than 1 kW/ton dry solids and is hence not considered here. In Figure 4-4, the energy requirement of HTHP pretreatment as a function of dry solids content (bottom x-axis) and operating temperature is shown. The

thermal energy yield from anaerobic digestion as a function of volatile solids destruction is shown on the secondary top x-axis. Both the gross thermal energy yield and the yield of high grade heat, assuming an efficiency of conversion of 40%, is shown The intersections of the high grade heat output with the energy requirement give the minimum attainable volatile solids and maximum dry solids content which must be achieved in order to make the pretreatment process energetically favourable. The difference between operating lines to the right of this point give the nett high grade heat generated. As stated in the assumptions, no heat recovery is included in the calculation. However, with heat recovery from the heated pretreated stream to the incoming stream can reduce the energy requirements effectively.

$$\hat{E}_{TH} = \frac{1}{DS} \cdot Cp_{slurry} \cdot \Delta T$$ **Equation 4-1**

where $\hat{E}_{TH}$ = energy required for thermal hydrolysis [W/kg]
DS = dry solids content (%)
Cp_{slurry} = heat capacity of slurry [J/kg.K]
ΔT = temperature change [K]

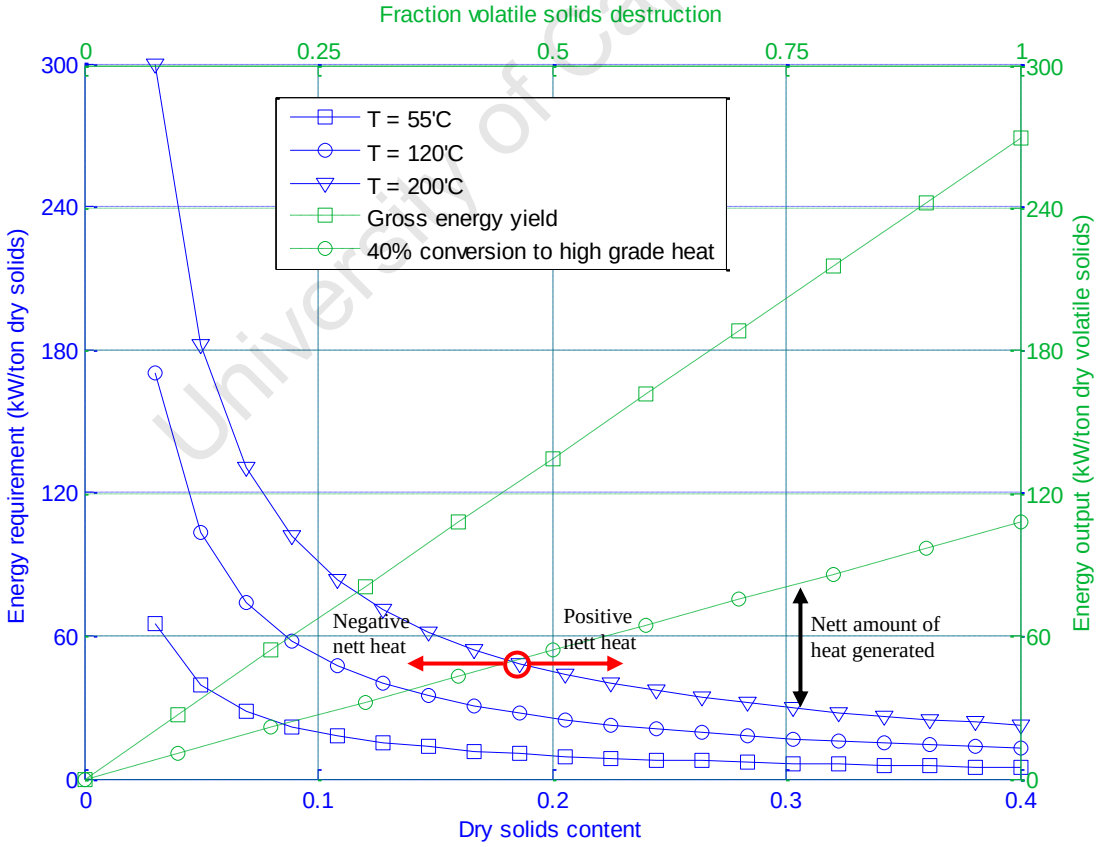


Figure 4-4: Energy requirement for thermal hydrolysis

(2) Energy requirements of high pressure homogenisation

The energy requirement from high pressure homogenisation may be calculated from Equation 4-2. The following assumptions were made in the calculation:

- The slurry density was assumed to be the same as water (1000 kg/m^3)
- A single pass through the homogeniser was used

The energy requirement for operating pressures of 30, 50 and 100 MPa are shown in Figure 4-5. It is clear that this is orders of magnitude greater than the energy output of the digester. Thus, lower operating pressures, viz., 8, 9 and 10 MPa, are shown in Figure 4-6 (blue axis) with the energy yield from anaerobic digestion given on the secondary axis (green axis).The same analysis used for thermal hydrolysis can be followed. From this analysis it is clear that high pressure homogenisation for treating sludge to improve digestibility is not feasible at the very high pressures used in some studies. Furthermore, these calculations assume only one pass. The energy requirement will increase proportionally to the number of passes. As the operating pressure of the Microsludge system is not quoted, it is difficult to tell whether this system is energy positive. The manufacturers claim so, thus is must operate at a pressure of less than 10 MPa and a dry solid content of greater than 12% (as highlighted by point A on Figure 4-6). The process would have to achieve a volatile solids content of greater than 35%, which according to literature is possible.

$$\hat{E}_{HPH} = \frac{1}{DS} \cdot \frac{1}{\rho_{slurry}} \cdot P$$ **Equation 4-2**

where $\hat{E}_{HPH}$ = energy required for high pressure homogenisation [W/kg]
DS = dry solids content (%)
ρ_{slurry} = slurry density [kg/m^3]
P = operating pressure of homogeniser [Pa]

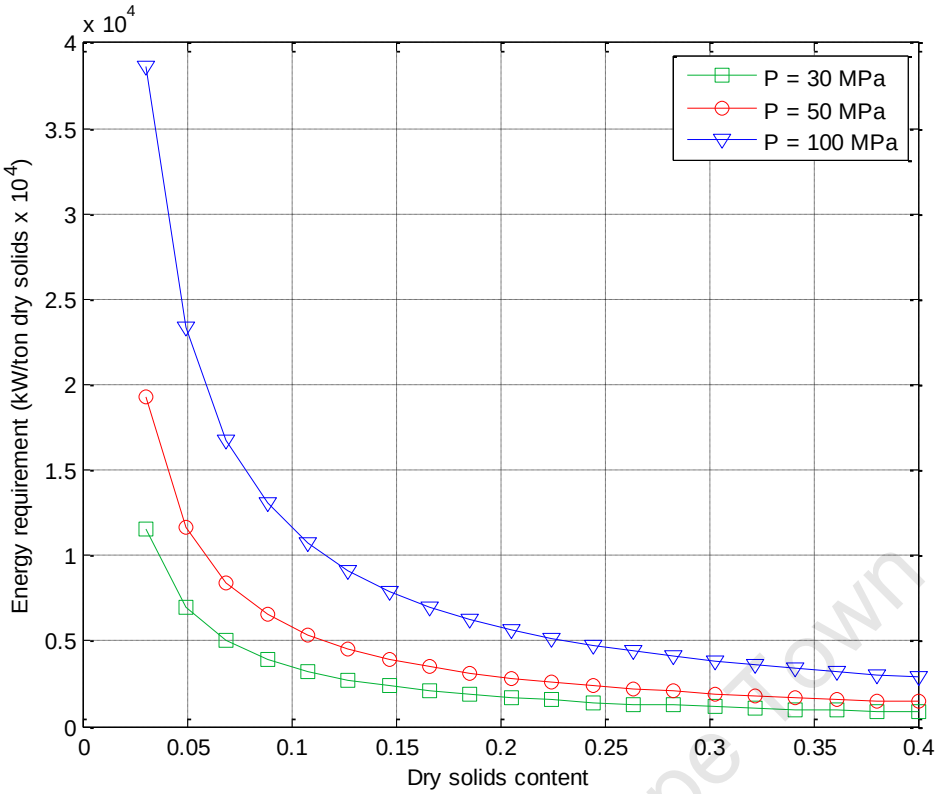


Figure 4-5: Energy requirements for high pressure homogenisation (high pressure range)

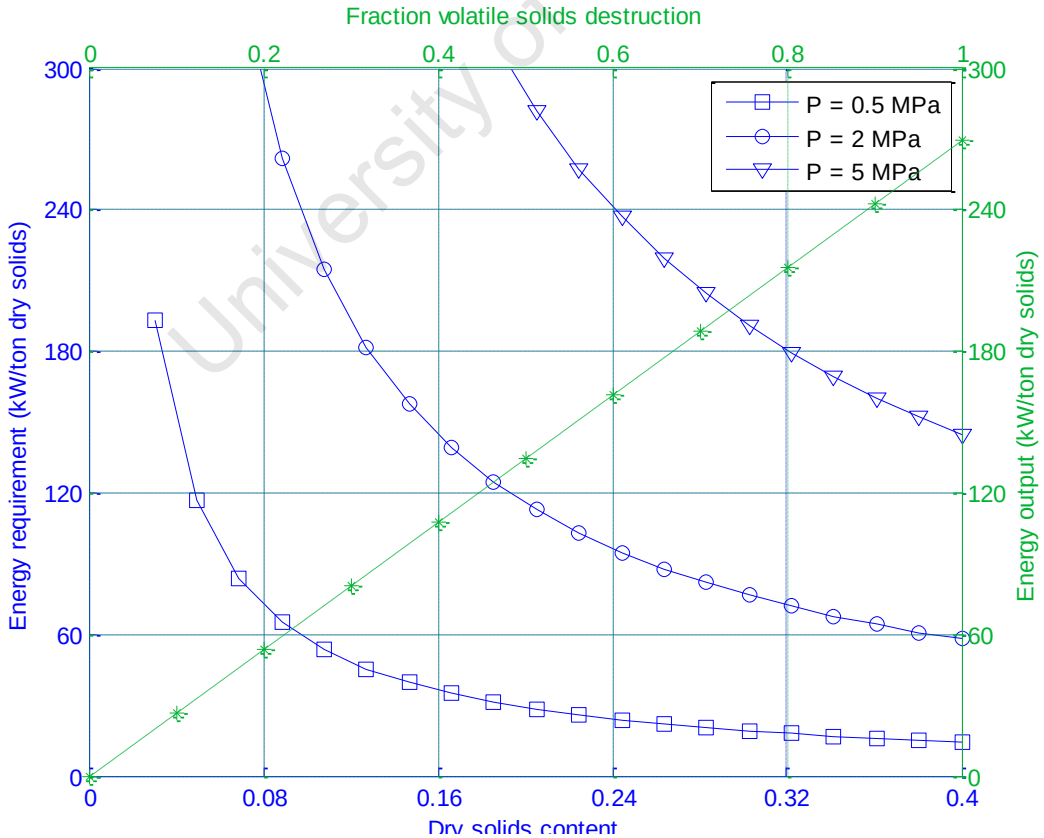


Figure 4-6: Energy requirement for high pressure homogenisation (low pressure range)

4.1.3. Conclusions

The complex nature of biogenic waste streams means that biodegradability should ideally be tested experimentally. Some idea about the maximum extent of biodegradability can be inferred from lignin content of the fuel. Representative volatile solids destruction can be obtained from the Biochemical Methane Potential assay.

Anaerobic digestion pretreatments such as thermal hydrolysis and high pressure homogenisation can be used increase the biodegradability of microbial sludges by disrupting the cell so that the intracellular proteins and carbohydrate storage products are released. The energy requirements of high pressure homogenisation at pressures of 30 to 50 MPa exclude its use for improving digestibility, as a positive nett energy yield is not possible. HTHP pretreatments are the recommended pretreatments as the analysis showed a positive nett energy yield at dry solids contents greater than 20% and at volatile solids destruction of greater than 50%, assuming no heat recovery and a thermal efficency of 40%. Heat recovery from the heated pretreated stream reduces the energy requirements and hence makes the pretreatments viable at lower volatile solids destruction.

4.2. The effect of volatile combustion in fluidised beds

4.2.1. Introduction

In order to optimise energy recovery from biomass combustion, control of overbed burning is necessary. The objective of this section is to investigate the effect of temperature on overbed burning and combustion efficiency experimentally for biogenic materials, using wood chips and sewerage sludge as examples. Here the combustion efficiency is defined as the actual heat released into the bed as a fraction of the maximum possible heat generation from the fuel.

A commonly cited industrial problem of combusting biogenic fuels in bubbling fluidised beds is overbed burning which can lead to runaway freeboard temperatures and difficulties in maintaining the bed temperature (North, 2005; Fountain, 2008 and Werther and Ogada, 1999). Overbed burning occurs when the volatiles released by the solid fuel burn in the freeboard above the bed. Biogenic fuels are characterised by their high volatile content and are hence more susceptible to overbed burning. Previous experimental work on bubbling

fluidised beds by Turnbull and Davidson (1984), Hesketh and Davidson (1991) and Ogada and Werther (1996) shows that the location of combustion of the gas or volatile component is dependant on the bed temperature, with higher bed temperatures promoting in-bed bubble phase combustion. The location of volatile component release and combustion is expected to be different for overbed and underbed feeding (Werther and Ogada, 1999), as underbed feeding promotes a longer residence time in the bed.

4.2.2. Factors affecting volatile combustion

4.2.2.1. Hydrocarbon gases and coal volatiles

When fluidising a bed of sand with a stoichiometric mixture of propane and air (used to model volatiles), the ignition of the fuel and location of combustion is dependent on the bed temperature. At low bed temperatures (<600˚C), an ignition source is needed to burn the gas in the freeboard. At temperatures around 750˚C, ignition is spontaneous in the freeboard. At bed temperatures higher than some critical temperature (T_{crit1}), estimated as 770°C (Turnbull and Davidson, 1984), the gas burns in the bed (van der Vaart, 1988). Turnbull and Davidson (1984) presented three regimes of gas combustion to formalise these observations

- Regime A, where $T_{bed} < T_{crit1}$ and gases do no burn in the bed
- Regime B, where $T_{crit1} < T_{bed} < T_{crit2}$ and gases burn only in the bubble phase
- Regime C, where $T_{bed} > T_{crit2}$ where combustion occurs both in the bubble and particulate phase.

T_{crit1} can be determined by visual inspection of the freeboard or by measuring pressure fluctuations in the freeboard (van der Vaart, 1988). The difference between T_{crit1} and T_{crit2} is estimated to be 120 K (Turnbull and Davidson, 1984).

Dennis *et al.* (cited in Turnbull and Davidson, 1984) proposed that particulate phase combustion was inhibited by sand particles quenching the radical reaction at the surface of the bubble and that higher bed temperature promoted the reaction. Van der Vaart (1988) investigated the effect of bed height and fluidisation velocity on T_{crit}, and found that increasing bed height and lowering fluidisation velocities decreased T_{crit}. Both these conditions permit a longer residence time in the bed. From these observations van der Vaart (1998) proposed that the combustion reaction was inhibited by an ignition delay. If the gas residence time was long enough, ignition would occur in the bed. Howevr, later work

supports the theory of the radical reaction being quenched by the presence of solid particles (Hayhurst and Parmar, 1998).

In models relating to the combustion of coal in a fluidised bed, assumptions must be made about the location and release of the volatile content. Three modelling approaches have been developed (LaNauze cited in Werther and Ogada, 1995):

- The release and combustion of the volatiles takes place instantaneously at the feed point, *i.e.* the rate of devolatilisation is much greater than the rate of mixing and the rise time of the particle;
- The release and combustion of volatiles occur uniformly throughout the bed, *i.e.* the rate of devolatilisation is less than the rate of mixing and similar to the rise time of the particle through the bed;
- The volatiles are released as a plume which does not mix with the oxygen and so only freeboard combustion occurs.

For coal particles, devolatilisation times are of the order of 10 to 20s, and 70% of devolatilisation occurs within 25 to 30% of the time for total devolatilisation. Thus, 2 to 7s is the devolatilisation time which should be used when determining the location of release of volatiles. The average rise velocity of a coal particle in a bed can be calculated from Nienow (1978) (cited in Turnbull and Davidson, 1984), according to Equation 4-3, with typical values ranging from 2 to 10 s. Thus the average rise time of the fuel particles is similar to the devolatilisation time and combustion should occur uniformly throughout the bed.

$$U_R = 0.15 \cdot (U - U_{mf})^{0.5} \qquad \textbf{Equation 4-3}$$

where U_R = average rise velocity of coal particle in bed [m/s]
U = superficial velocity [m/s]
U_{mf} = minimum fluidisation velocity [m/s]

An experiment by Hesketh and Davidson (1991) used char and a mixture of propane and air to model the fixed carbon and volatile content of coal respectively. The burnout time for a batch of char added to a hot fluidised bed can be calculated by carrying out an energy balance over the bed, as given by Hesketh and Davidson (1991) and presented in Equation 4-4.

$$M_{bed} \cdot Cp_{bed} \cdot \frac{dT_{bed}}{dt} = \Delta H_c \cdot r_c \cdot N_p + h \cdot A_h \cdot (T_{furnace} - T_{bed}) - m_{gas} \cdot Cp_{gas} \cdot (T_{bed} - T_{inlet}) \quad \textbf{Equation 4-4}$$

where M_{bed} = mass of bed material [kg]
Cp_{bed} = heat capacity of bed material [J/kg.K]
A_h = heat transfer area [m^2]
T_{bed} = bed temperature [K]
$T_{furnace}$ = furnace temperature [K]
ΔH_c = heat of combustion [J/kmol]
m_{gas} = mass flowrate of gas [kg/s]
r_c = rate of combustion [kmol/particle/s]
Cp_{gas} = heat capacity of gas [J/kg.K]
h = overall furnace to bed heat transfer coefficient [$W/m^2.K$]
T_{inlet} = inlet temperature [K]
N_p = number of particles

After burnout of the char, the bed temperature declines to its previous steady temperature according to Equation 4-4 with r_c set to zero. Identification of the point on the burnout curve where this relationship holds gives the burnout time of the char. It was found that the burnout time for the char increased at higher temperatures when the bed was fluidised with the air/propane mixture as opposed to air alone. The effect was attributed to propane combustion proceeding in the bed at higher temperatures, thus consuming relatively more oxygen than at low temperatures. The consumption of oxygen by the volatiles decreased the oxygen available for combustion of the char and increased the burnout time of the char.

Turnbull (1983) investigated the fluidised bed combustion of volatiles from coal by carrying out an energy balance over the freeboard. It was found that up to 70% of the energy released in the freeboard was returned to the bed by the solids falling back into the bed. Turnbull used Pemberton's model (1982) to predict the flux of solids into the freeboard. Pemberton assumed that a single layer of particles carried on the nose of a bursting bubble was projected into the freeboard. The high heat transfer between bed and freeboard predicted by Turnbull (1983) is contrary to industrial experience and other experimental work. The anomaly is explained by "ghost bubbles", fuel rich bubbles which pass through the splash zone and burn above the splash zone where the heat released is not returned to the bed (Turnbull, 1983).

4.2.2.2. Biomass fuels

Biomass fuels are characterised by their high volatile content. In the case of the fluidised bed combustion of sewerage sludge, up to 80% of the sludge carbon is released as volatiles. Comparisons of the oxygen profile through a fluidised bed combusting sewerage sludge showed that it was very close to that of a bed combusting hydrocarbon gas (Ogada and Werther, 1995). Critical issues associated with the combustion of these types of fuel include separation of volatile and fixed carbon content in the combustor and particle attrition and fragmentation (Chirone *et al.*, 2008). This can lead to a loss in carbon conversion as the

material is lost from the reactor unconverted, or to high freeboard temperatures where the volatiles or fines combust in the freeboard.

Conflicting results appear in the literature regarding the location of the release of volatile components. Ogada and Werther (1996) found that the time for devolatilisation of a single particle of sewerage sludge was much faster than the solids mixing time, and as a result uniform combustion should take place in the bed. However, in-bed combustion does not explain the high freeboard temperatures often seen in industrial units (North et al, 2005; Fountain, 2008 and Werther and Ogada, 1999). High freeboard temperatures could be due to volatiles leaving the bed in bubbles and burning only in the freeboard, or lack of segregation of sludge particles within the bed, which could result in local oxygen deficient regions and associated release of CO to the freeboard. Ogada and Werther (1996) analysed C_xH_y profiles (representative of volatiles) in a reactor, and detected low concentrations just above the feed point, with higher concentrations seen in the splash zone and freeboard. Regarding particle segregation, some researchers found that the action of a high velocity jet broke up the fuel particles into smaller pieces (Ruoss (1991) cited in Werther and Ogada, 1999) whereas others have seen that the fuel particles retain their shape. Work by Chirone *et al.* (2008) show an intact ash skeleton left behind from sludge combustion which indicates a lack of segregation of particles.

Moisture content and particle size of the fuel also affects the location of release of volatiles and their combustion. Work done by North *et al.* (2005) showed that the loss in heat from the co-firing of a sludge of coffee grounds and coal was approximately 7% of total fuel input, which they attributed to overbed combustion of the coffee sludge. To overcome this, the fuel was partially dried by introducing it at a high level in the freeboard. Borodulya (cited in Werther and Ogada, 1999) noted that sludge particles dropped onto the top of an 800°C bed remained intact and floated at the top before submerging into the bed for a short period. The larger, moist particle burnt at the surface of the bed with a long and sooty flame whereas particles smaller than 10 mm exhibited no flame, indicating combustion within the bed.

4.2.3. Experimental materials and methods

4.2.3.1. Fuels

The following fuels were investigated in this study:

- Woodchips: Catsan Natural Wood cat litter was sieved to provide a size range 2814 μm to 4000 μm. The cat litter is made from crushed briquetted softwood sawdust and was chosen because of ease of feeding the spherical particles into the combustor
- Sewerage sludge: Thames Water supplied samples of air-dried sewerage sludge in an extruded form. The sludge was crushed and sieved to provide a size range 2814 μm to 4000 μm.

CHN elemental composition, determined by microanalysis, is given in Table 4-4. Oxygen content was calculated by difference. The ash content was determined gravimetrically by heating the sample to 550°C for 5 hours. The moisture content was determined gravimetrically by heating the sample to 120°C until constant mass was achieved.

Table 4-4: CHNO analysis of wood and sewerage sludge

	Wood	Sludge
C	46.6%	46.8%
H	6.04%	6.60%
O	34.3%	25.1%
N	0.00%	4.94%
Ash	13.1%	16.6%
Moisture	2.55%	1.88%

4.2.3.2. Experimental setup

The reactor used in this work was a 112 mm high 316 stainless steel tube with an internal diameter of 77 mm. The distributor plate was mounted 500 mm above the bottom of the reactor and had thirty-seven 0.4 mm holes arranged in a triangular pattern. Air entered the plenum chamber via a 6.35 mm (ID) stainless steel pipe. A diagram of the setup is shown in Figure 4-7.

Aluminium oxide with a size fraction of +297-350 μm was used as the bed material. Aluminium oxide was selected rather than silica sand, since silica sand was ground down to a very fine dust in the feedarm producing a potentially hazardous crystalline dust. The composition of the bed material is given in Table 4-5.

Table 4-5: Bed material composition

Al_2O_3	95.2%
TiO_2	2.9%
Fe_2O_3	0.2%
CaO – MgO	0.3%
SiO_2	1.2%

Heat was provided by six Crusilite silicon carbide heating rods (Kanthal Ltd) placed on either side of the reactor tube at heights of 250 mm, 480 mm and 710 mm from the reactor base. The lower two rods provided most of the heat to the bed, whilst the top four rods heated the air passing through the freeboard to a temperature higher than the bed temperature. The rods drew a current of 22 A from the mains supply (240 V) to give a range of furnace power of 0 to 5300 W. In these experiments, the power was set to a constant value using a thyristor controller.

The reactor was insulated by a rectangular wall of firebricks 120 mm thick (JM36, Thermal Ceramics Ltd). The feed arm and bottom flange were also insulated. (Superwool Thermal Ceramics). The top of the furnace block was insulated with Superwool and ceramic board. The bottom feedarm was cooled by running water at ~0°C through copper coils to prevent the feed from devolatilising or partially combusting before entering the bed. The feedarm and cooling coils were insulated with Superwool.

The feeding system of the rig consisted of a feed hopper mounted on a brass block. A small helical doping screw, driven by a stepping motor, controlled the feedrate to the main auger. The main feed auger was water-cooled to prevent biomass from devolatilising or partially combusting in the feedarm. A positive pressure was maintained on the hopper at all times to prevent the backflow of sand into the feedarm or into the hopper.

The reliable operation of the hopper and doping screw system was challenging. Two operational problems arose regularly: the doping screw jammed or the feed formed a bridge in the hopper. The solution to the former was to use an optimal particle size of range 2814 μm to 4000 μm. The problem of the biomass bridging was ongoing and also found to have some dependence on particle size. A number of "anti-bridging" devices were crafted over the course of this work but no superior design emerged. It was also found that a helical screw feeder does not move the feed forward satisfactorily. It is suggested for future work that a machined auger be used.

4.2.3.3. Data acquisition

(1) Temperature measurements

The bed temperature was measured using a 3mm type-K thermocouple. A suction pyrometer was used to measure the freeboard temperature. A 1.5 mm K-type thermocouple was mounted

in a 6.35 mm steel pipe. A diaphragm pump drew air over the thermocouple at 4 m/s. This increased the Nusselt number from 2 to 8 and hence increased the convective heat transfer coefficient. The radiation shield and the high convective heat transfer ensured that the true gas temperature was measured, free from radiation effects.

(2) Flue gas analysis

The 6.35mm pipe used as the thermocouple radiation shield doubled as the sample line. The superficial velocity through the pipe was 4 m/s (vs. 0.4 m/s for the reactor). The offgas was drawn through a 200 μm glass microfibre filter to remove fine dust, tar and char. Two impinger flasks (one empty and one filled with glass beads) were placed in an ice bath to condense water vapour. Thereafter the gas was passed through a calcium chloride moisture trap. The concentrations of CO and CO_2 were measured continuously by an infrared gas analyzer (ABB Easyline I.R.) The O_2 concentration was measured by a Universal Exhaust Gas Oxygen (UEGO) sensor (Horiba MEXA 720). UEGO's are used primarily for monitoring air fuel ratios in motor vehicles. The gas stream is passed over a heated element to oxidise any remaining fuel, including char fines and tar. Hence, the decrease in O_2 concentration measured by the UEGO can be equated directly to the mols of carbon content fed to the reactor.

4.2.3.4. Experimental procedure

In order to prevent build up of ash in the reactor, 2.1 kg of fresh bed material was placed in the reactor tube at the start of each run. 30 to 45 g of fuel was placed in the hopper and the hopper lid was sealed. A purge of nitrogen of 4 l/min was passed through the hopper. The feed auger, was started to prevent sand flowing back into the feedarm. The air flowrate through the reactor was set such that U/U_{mf} was 7 to 7.5. An equivalence ratio (*i.e.* the ratio of mols fed to mols O_2 required for complete combustion) of approximately 2 for wood runs and 1.2 for sludge was used. The bed temperature was increased to the desired temperature by running the furnace at 50 to 80% power. Once at the desired temperature the power input was reduced slightly to maintain a constant bed temperature. At this stable temperature the fuel was fed into the bed by turning on the dosing screw. The offgas compositions, freeboard temperature and bed temperature were monitored and recorded.

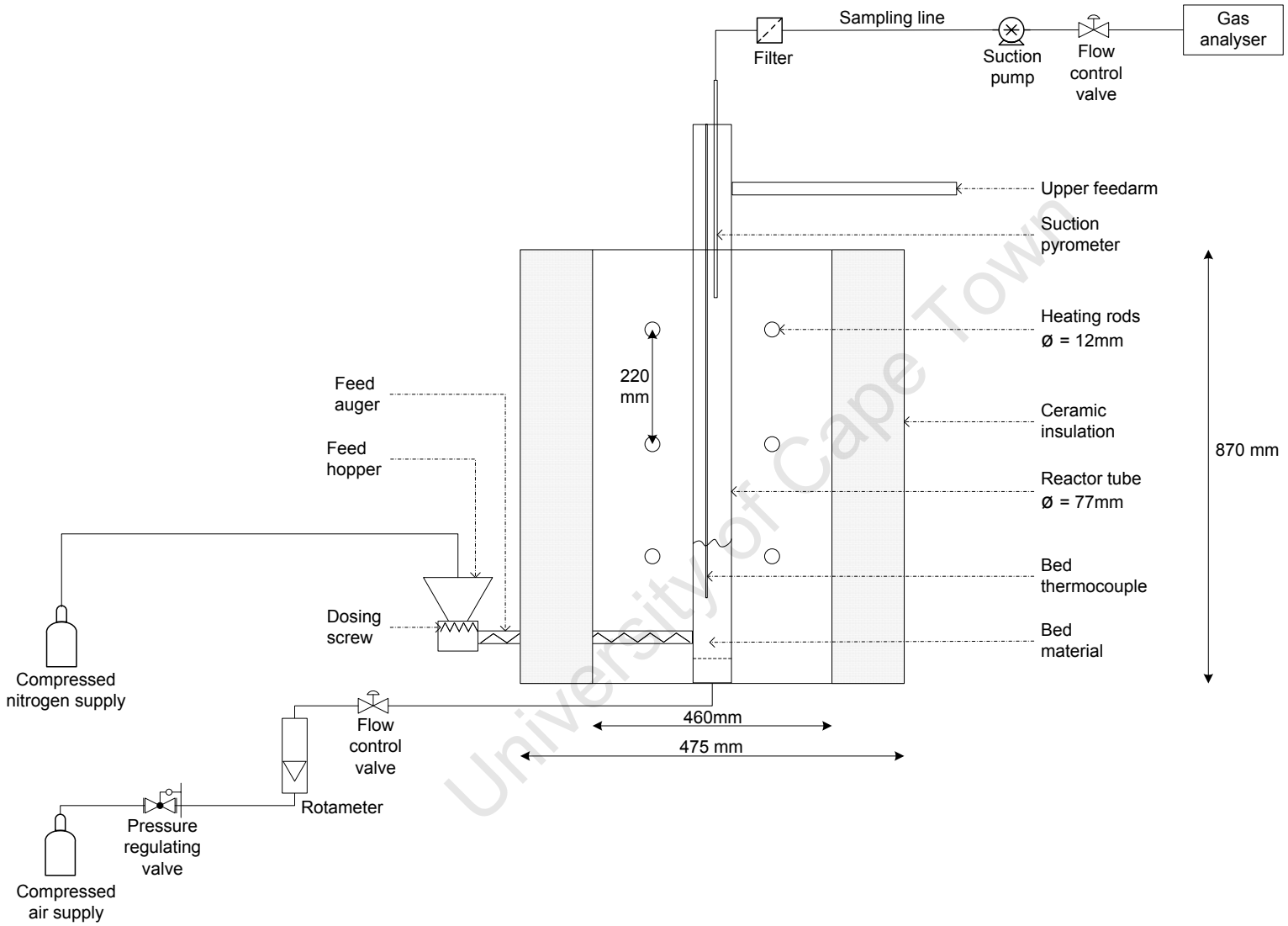


Figure 4-7: Diagram of reactor apparatus

4.2.4. Mass balances

4.2.4.1. Calculations

A carbon mass balance was carried out for the system. The total mols of carbon fed to the system in a particular run can be calculated from Equation 4-5. The total number of mols of carbon exiting the system as CO_2 and CO for a particular run can be calculated from Equation 4-6. The UEGO oxygen sensor measured the total oxygen consumed. As the flue gas passes over the sensor's heated element, any char fines and tar are oxidised. The total amount of oxygen consumed can be calculated from Equation 4-7. A sample plot of the raw data is given in Figure 4-8.

$$N_{C,in} = M_{fuel} \cdot \%C \cdot MM_C$$ **Equation 4-5**

where $N_{C,in}$ = carbon fed to the system [mol]
M_{fuel} = mass of fuel placed in hopper [kg]
%C = carbon content of fuel
MM_C = molar mass of carbon [kg/mol]

$$N_{C,out} = n_{gas} \cdot \int [CO_2]dt + \int [CO]dt$$ **Equation 4-6**

where $N_{C,out}$ = carbon exiting system in flue gas [mols]
n_{gas} = molar flow of gas through the reactor [mol/s]
$[CO_2]$ = molar concentration of CO_2 in the flue gas (as %)
[CO] = molar concentration of CO in the flue gas (as %)

$$N_{O2,out} = n_{gas} \cdot \int [O_2]dt$$ **Equation 4-7**

Where $N_{O2,out}$ = oxygen exiting system in flue gas [mols]
n_{gas} = molar flow of gas through the reactor [mol/s]
$[O_2]$ = molar concentration of CO_2 in the flue gas (as %)

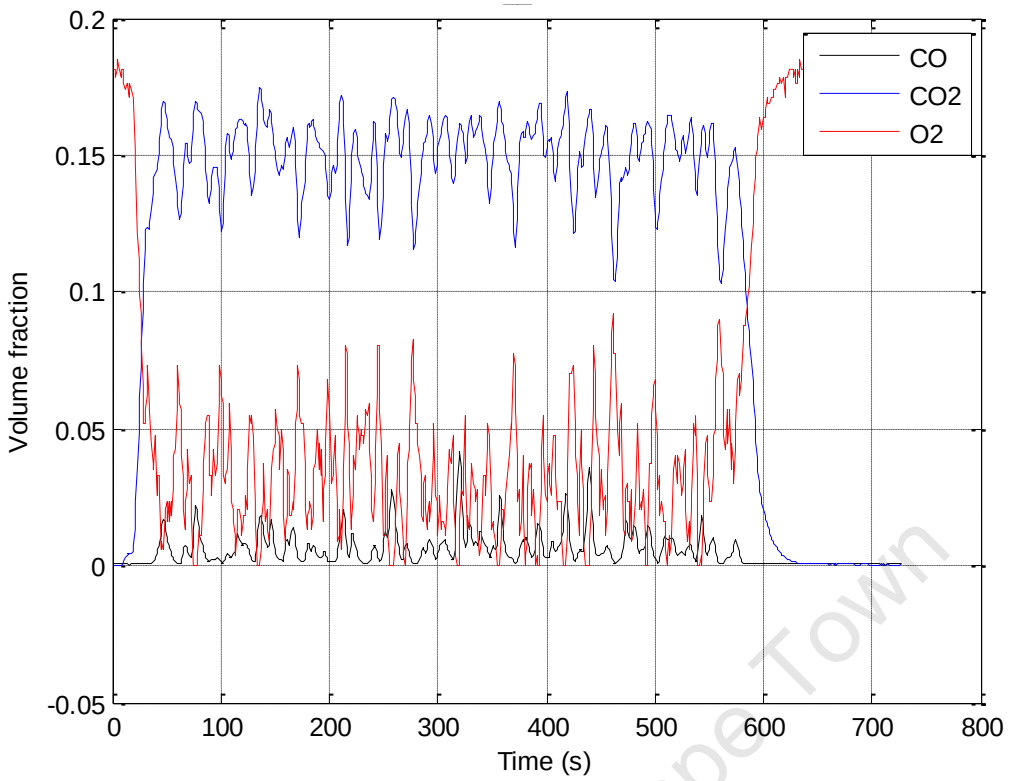


Figure 4-8: Sample plot of flue gas analysis (Run 16)

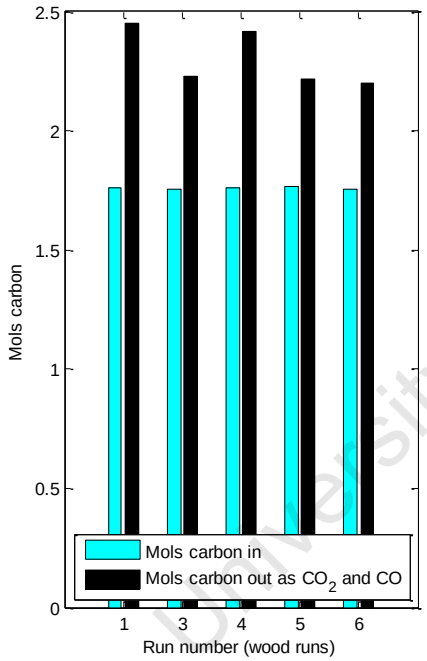
4.2.4.2. Results and discussion

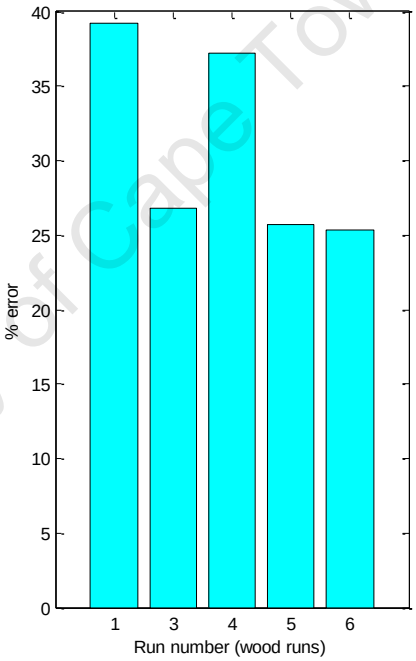
Figure 4-9 and Figure 4-10 give the carbon balance for the wood and sludge runs respectively and shows the implausible result of more mols of carbon exiting the system than entering the system. This could be due too a number of causes:

- Incorrectly calibrated instruments
- Incorrect value for carbon content of fuel
- Leak in sample line
- Leak in airflow line after the rotameter

The first two possibilities eliminated as the instruments were re-calibrated and the carbon content re-checked. A leak in the sample line would have diluted the CO_2 and CO in the flue gas stream which is not the case here. Our conclusion was that a leak in airflow line downstream of the rotameter resulted in a higher gas flowrate being measured than was being passed through the reactor. The perturbation energy balance over the bed (Equation 4-13) does not require knowledge of the gas flowrate. The energy balance over the freeboard does require the gas flowrate, but it is assumed that the error between the results remained constant and hence allows for their comparison.

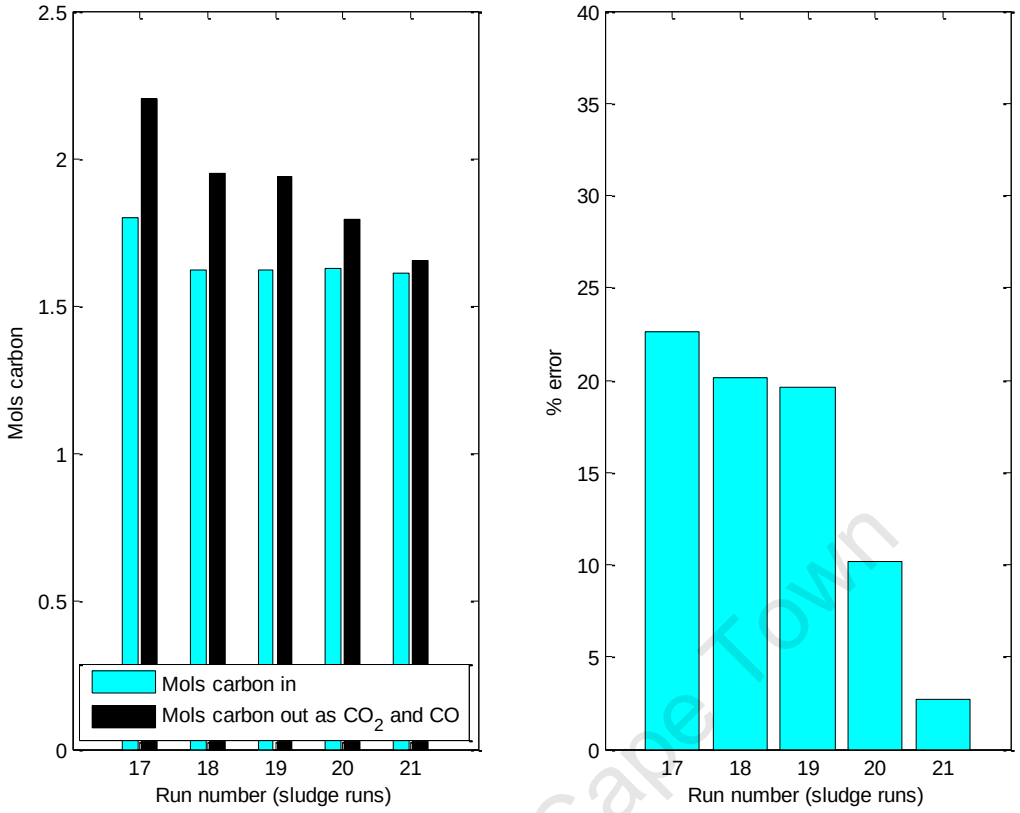
Figure 4-11 shows the mols of oxygen consumed, as measured by the UEGO sensor, and compares this to the mols of carbon released as CO_2 and CO. This is shown for selected runs where the UEGO sensor was available. As both these quantities are dependent on the molar flow of gas through the reactor, any error in the measurement of gas flow rate did not influence mass balance. The average error in the difference between oxygen consumed and carbon released as CO_2 and CO is 8%, which is within expected experimental error. This supports the conclusion that a leak in the airline is a likely reason for the carbon balance not holding.





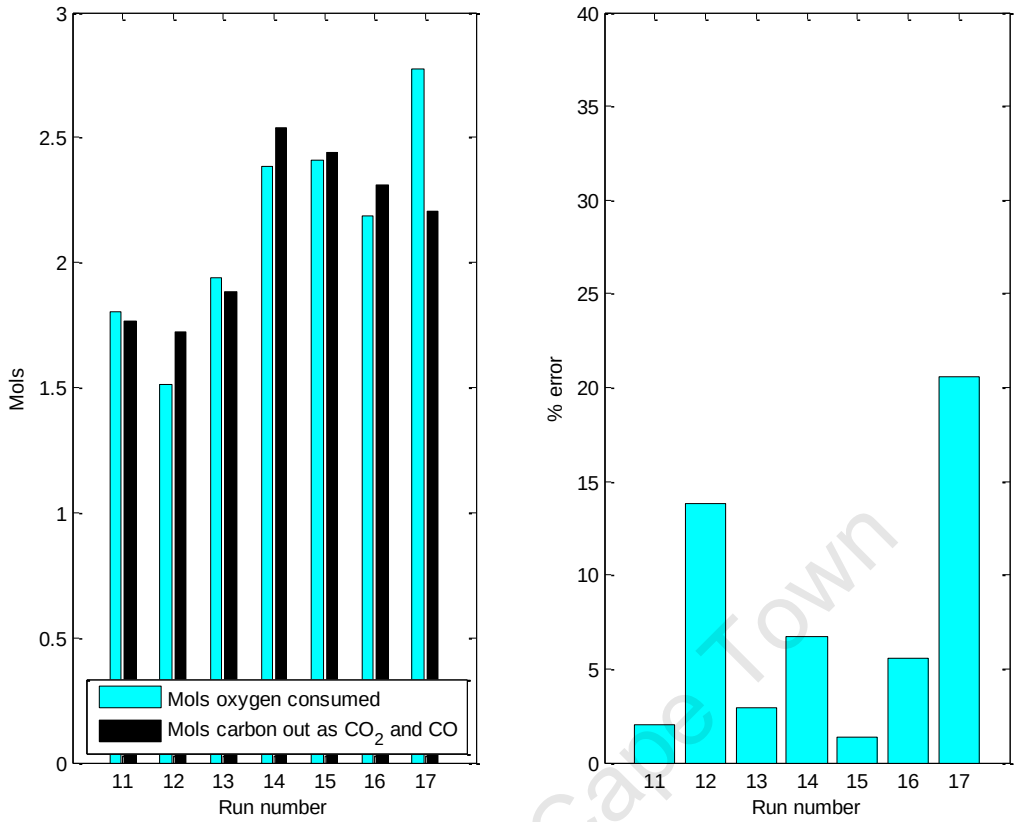
Run number	Run ID	Initial bed temperature (°C)	Fuel flowrate (g/min)	U/Umf	Equivalence ratio
1	'W__BC700'	714	2.2	6.5	2.0
3	'W__BC800'	805	2.4	7.7	1.9
4	'W__BC850'	866	2.5	8.4	1.8
5	'W__BC2__800'	805	2.0	7.1	2.0
6	'W__BC2__850'	870	2.0	7.4	1.9

Figure 4-9: Carbon balance for wood runs



Run number	Run ID	Initial bed temperature (°C)	Fuel flowrate (g/min)	U/Umf	Equivalence ratio
17	'S__750__O2'	760	2.3	6.2	1.5
18	'S__BC700'	715	3.8	6.7	1.1
19	'S__BC750'	759	3.3	6.9	1.2
20	'S__CB800'	800	3.8	7.0	1.0
21	'S__BC850'	848	3.2	7.2	1.1

Figure 4-10: Carbon balance for sludge runs



Run number	Run ID	Initial bed temperature (°C)	Fuel flowrate (g/min)	U/Umf	Equivalence ratio
11	'W__lowUlowER850'	838	2.9	4.8	0.8
12	'W__lowER__N2'	741	3.5	7.3	0.7
13	'W__lowER'	746	3.2	6.7	1.4
14	'W__OB750'	763	2.1	7.1	2.1
15	'W__OB850'	849	2.2	8.2	2.0

Figure 4-11: Mass balance for oxygen consumed and carbon released as CO_2 and CO for selected wood runs

4.2.5. Energy balance model

4.2.5.1. Model development

An energy balance for the system shown in Figure 4-12 can be described by Equation 4-8 prior to introducing the feed. At thermal steady state, Equation 4-8 reduces to Equation 4-9. Excluding the possibility of overbed burning, the energy balance over the bed during feeding of fuel is given by Equation 4-10, ,where Q_{in} is of the form of Equation 4-11. The heat input did not change before and after feeding commenced. With the assumption that for a short time period after feeding starts the heat loss from the system remains unchanged, Equation 4-9 can

be substituted into Equation 4-10 to give Equation 4-12. Equation 4-12 is expressed in deviation variables, *i.e.* $T^* = T_{bed}-T_{bed,ss}$. As feeding starts (*i.e.* t = 0, $T_{bed} = T_{bed,\ ss}$ and $T^* = 0$), Equation 4-12 reduces to Equation 4-13.

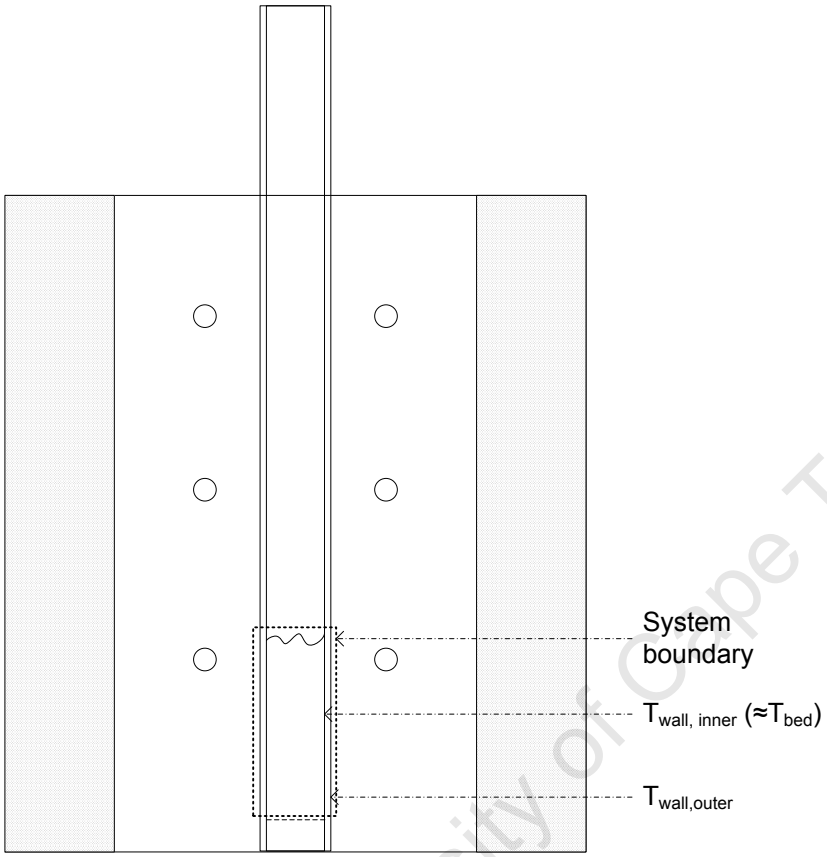


Figure 4-12: Reactor showing system boundary

$$Cp_{bed} \cdot M_{bed} \cdot \frac{dT_{bed}}{dt} (= 0) = Q_{in} - \dot{m}_{air} \cdot Cp_{air} \cdot (T_{bed}(t) - T_{inlet}) \qquad \textbf{Equation 4-8}$$

Condition: No fuel fed; unsteady temperature
Where Cp_{bed} = heat capacity of bed [J/kg.K]
M_{bed} = mass of bed material [kg]
$T_{bed,ss}$ = steady state bed temperature [K]
Q_{in} = heat input to bed from furnace [W]
Cp_{air} = heat capacity of gas [kJ/kg.K]
T_{inlet} = inlet air temperature [K]
m_{air} = mass of gas flowrate through bed [kg/s]

$$Q_{in} = \dot{m}_{air} \cdot Cp_{air} \cdot (T_{bed,ss} - T_{inlet}) \qquad \textbf{Equation 4-9}$$

Condition: no fuel fed; steady temperature

$$Cp_{bed} \cdot M_{bed} \cdot \frac{dT_{bed}}{dt} = Q_{in} - \dot{m}_{air} \cdot Cp_{air} \cdot (T_{bed}(t) - T_{inlet}) + m_{biomass} \cdot \Delta H_c$$ **Equation 4-10**

Condition: Fuel fed; unsteady temperature
Where $m_{biomass}$ = mass flowrate of biomass [kg/s]
ΔH_c = heat of combustion of fuel [J/kg]
T_{bed} = bed temperature [K]

$$Q_{in} = U_h \cdot A_h \cdot (T_{wall,outer} - T_{bed})$$ **Equation 4-11**

where Q_{in} = heat input to bed [W]
U_h = overall heat transfer coefficient [W/m^2.K]
A_h = heat transfer area [m^2]
$T_{wall,outer}$ = outer reactor wall temperature [K]

$$Cp_{bed} \cdot M_{bed} \cdot \frac{dT^*}{dt} = -\dot{m}_{air} \cdot Cp_{air} \cdot T^*(t) + m_{biomass} \cdot \Delta H_c$$ **Equation 4-12**

Condition: fuel fed; deviation variable used
where $T^* = T_{bed} - T_{bed,ss}$ [K]

$$Cp_{bed} \cdot M_{bed} \cdot \frac{dT^*(t=0)}{dt} = m_{biomass} \cdot \Delta H_c$$ **Equation 4-13**

Condition: fuel fed; time = 0

4.2.5.2. Analysis of experimental data

The analytical solution to Equation 4-12 given by Equation 4-14. The constants A and α were found by fitting a curve of the form of Equation 4-14 to the bed temperature experimental data, using a non-linear regression routine in Matlab. The code is given in Appendix A.5 and the experimental data with the fitted curves are plotted in Appendix A.3. A sample plot is shown in Figure 4-13.

As feeding starts (*i.e.* at t = 0), Equation 4-14 reduces to **Error! Reference source not found.**.

$$T^*(t) = A \cdot (1 - e^{-\alpha \cdot t})$$ **Equation 4-14**

Where $T^* = T_{bed} - T_{bed,ss}$

To simplify the analysis, the point of interest is the perturbation point; the point at which feeding commences *i.e.* Equation 4-13. The right hand side of Equation 4-13 gives the potential maximum amount of heat released to the bed. The left hand side of Equation 4-13 is

obtained from the experiment work as $\frac{dT(t=0)}{dt} = A \cdot \alpha$. This allowed the combustion efficiency to be calculated as in Equation 4-15.

$$\eta = \frac{Q_{experimental}}{Q_{maximum}} = \frac{\frac{dT(t=0)}{dt}}{m_{biomass} \cdot \Delta H_c} = \frac{A \cdot \alpha}{m_{biomass} \cdot \Delta H_c}$$ **Equation 4-15**

where η = combustion efficiency
$Q_{experimental}$ = experimentally observed heat release to the bed
$Q_{maximum}$ = maximum possible heat release to the bed

The contribution of overbed burning to any loss in heat input to the bed could be calculated by the initial jump in freeboard temperature, as seen in Figure 4-13 and according to Equation 4-16.

$$Q_{fb} = \dot{m}_{gas} \cdot Cp_{gas} \cdot \Delta T_{fb}$$ **Equation 4-16**

where Q_{fb} = heat lost due to overbed burning (W)
m_{gas} = mass flow of gas through the bed (kg/s)
Cp_{gas} = heat capacity of gas (taken to be air) (kJ/kg.K)
ΔT_{fb} = jump in freeboard temperature (K)

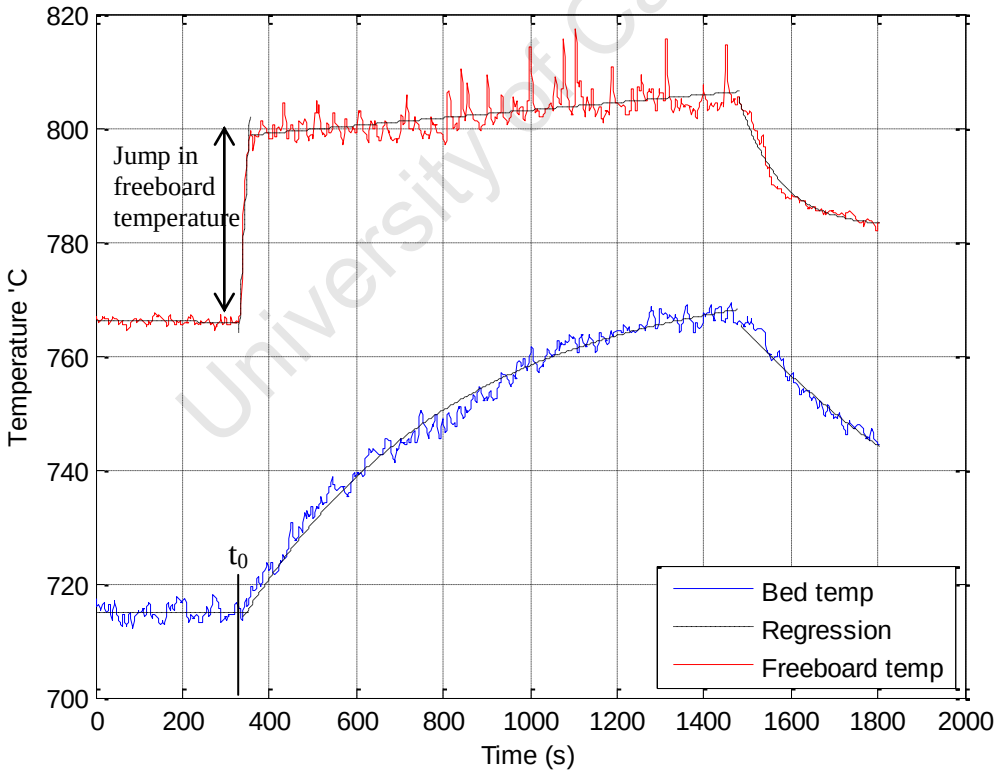


Figure 4-13: Sample plot of raw data (Run: 1)

4.2.5.3. Discussion on energy balance model

(1) Calculation of Q_{in}

Without knowledge of $T_{wall,\ outer}$, Q_{in} cannot be measured directly. However, if it is shown that the heat loss though the firebricks (given by Equation 4-17) is negligible, then the electrical heat input from the heating rods (given by Equation 4-18) could be used to determine Q_{in}. The scheme shown by Figure 4-14 was used to determine the inputs to $Q_{loss.}$ Using this approach, Q_{loss} was estimated to be 2400 W. This is significant compared to the 3000 W input from the electrical furnace. Hence, the heat input from the furnace could not be used to calculate Q_{in}, as explained schematically by Figure 4-15. Thus, the best method of analysis for the system is that given by the perturbation equation, Equation 4-13. All calculations are shown in Appendix A.3.2.

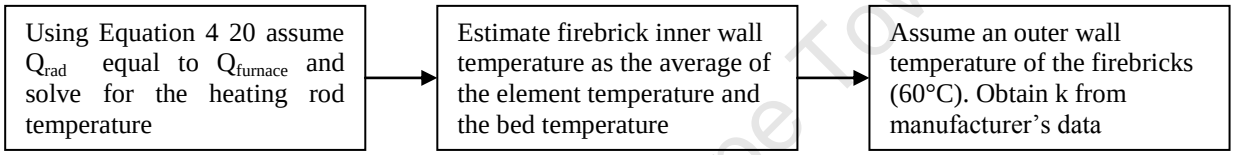


Figure 4-14: Determining Q_{loss}

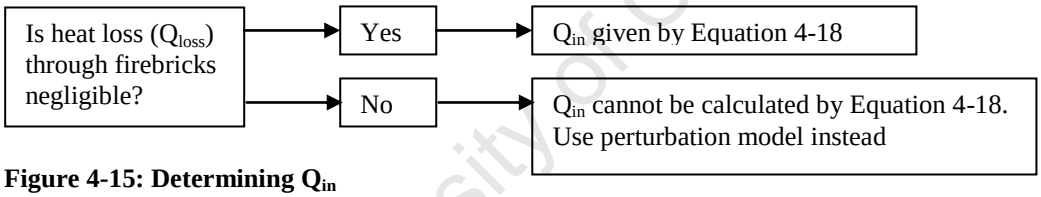


Figure 4-15: Determining Q_{in}

$$Q_{loss} = \frac{k}{l} \cdot A_h \cdot \left(T_{wall,inner} - T_{wall,outer} \right) \qquad \textbf{Equation 4-17}$$

where Q_{loss} = heat loss from bed (W)
k = conductivity of the firebricks (W/m.K)
l = thickness of firebricks (m)
$T_{wall,inner}$ = inner wall temperature of firebrick (K)
$T_{wall,\ outer}$ = outer wall temperature of firebricks (K)

$$Q_{furnace} = \%power \cdot I \cdot V \qquad \textbf{Equation 4-18}$$

where $Q_{furnace}$ = heat input to bed from furnace (W)
%power = percentage power input from furnace (~55%)
I = current through furnace heating rods (amps)
V = potential difference across heating rods (V)

$$Q_{radiative} = A_{hr} \cdot \sigma \cdot \varepsilon \cdot T_{element}^4$$ **Equation 4-19**

where $Q_{radiative}$ = radiative heat input from furnace (W)
ε = emissivity of heating rods (= 0.7)
A_{hr} = radiative heat transfer surface (m^2)
σ = Stefan Boltzmann constant ($5.67 \cdot 10^{-8}$ $W/m^2/K^4$)
$T_{element}$ = element temperature (K)

(2) Rate of combustion

The rate of combustion of biogenic fuels such as wood and sewerage sludge is very fast. This is seen in Figure 4-8 from the instantaneous jump in CO_2 concentration on commencement of feeding fuel, and the immediate drop in CO_2 concentration at the end of the run. Thus, the rate of combustion is controlled by the rate of feeding.

(3) Thermal mass of the bed

The term $Cp_{bed} \cdot M_{bed}$ includes the thermal mass of the steel wall in contact with the bed. This is justified by considering the high heat transfer coefficient between the bed and the wall, which ranges from 200 to 500 $W/m^2/K$ for bubbling fluidised beds and the Biot number of the bed-wall system. The Biot number of the bed-wall system was 0.1, meaning that the temperature profile through the wall is constant. Calculations are shown in Appendix A.3.2.

(4) Model limitations

Equation 4-14 is only used to model the initial section of the run, allowing for the complex heat transfer system to be simplified.

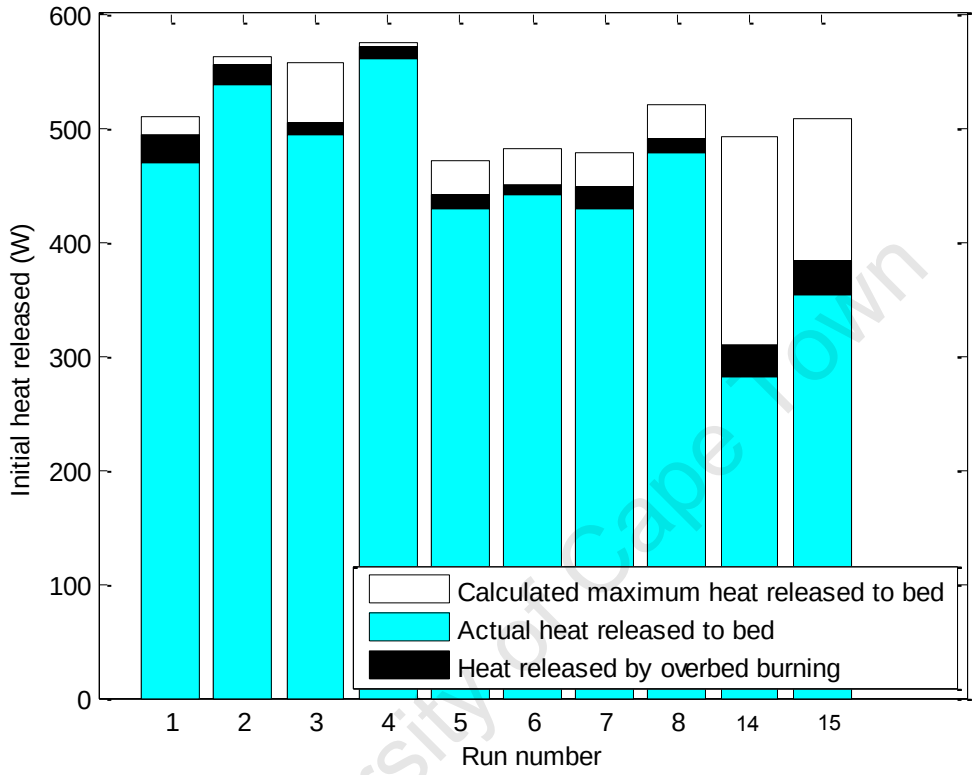
4.2.6. Results

Comparing the energy balance model derived in the Section 4.2.5 and to the experimental data, the combustion efficiency at different temperatures was investigated. The effect of underbed and overbed feeding was also considered.

4.2.6.1. Underbed and overbed feeding of wood

Figure 4-16 shows the maximum amount of heat released to the bed as feeding commences calculated as $m_{biomass} \cdot \Delta H_c$. This is compared to the experimentally observed heat release, $(Cp_{bed} \cdot M_{bed}) \frac{dT(t=0)}{dt}$ where $\frac{dT(t=0)}{dt}$ was obtained from the regression analysis of the experimental data. The portion of the heat lost to overbed combustion is also shown. The operating conditions of each run are given. The maximum potential heat release ranges from

470 to 570 W depending on the fuel flowrate. The experimentally observed heat release to the bed ranges from 430 to 560 W. The contribution from overbed burning is on average 14 W for the underbed feeding and 28 W for overbed feeding which is less than 10% of the calculated maximum heat release. The unaccounted for heat loss in this these runs ranges from 3 to 52 W for underbed feeding and from 124 W to 184 W for overbed feeding.



Run number	Feed location	Initial bed temperature (°C)	Fuel flowrate (g/min)	U/Umf	ER
1	Underbed feeding	714	2.2	6.5	2.0
2	Underbed feeding	758	2.4	7.0	1.9
3	Underbed feeding	805	2.4	7.7	1.9
4	Underbed feeding	866	2.5	8.4	1.8
5	Underbed feeding	805	2.0	7.1	2.0
6	Underbed feeding	870	2.1	7.4	1.9
7	Underbed feeding	709	2.1	6.4	2.1
8	Underbed feeding	750	2.2	7.0	2.0
14	Overbed feeding	763	2.1	7.1	2.1
15	Overbed feeding	849	2.2	8.2	2.0

Figure 4-16: Partition of heat release between bed and freeboard for wood runs

To display the results independent of fuel feedrate, the combustion efficency as a function of initial bed temperature is shown in Figure 4-17. The 95% confidence intervals displayed are acquired from the regression analysis *i.e.*it is a mathematical error not an experimental error.

The experimental error is not shown explicitly but an idea can be gauged from the results repeated at similar temperatures. Figure 4-17 shows that for all wood underbed-fed runs, approximately 90% of the heat is released to the bed. There is no clear dependence of combustion efficiency on temperature for these runs. From Figure 4-17 it can be seen that the combustion efficiency of overbed feeding is ~65 %. Some dependence on temperature is evident as the combustion efficiency increases from 750° to 850°.

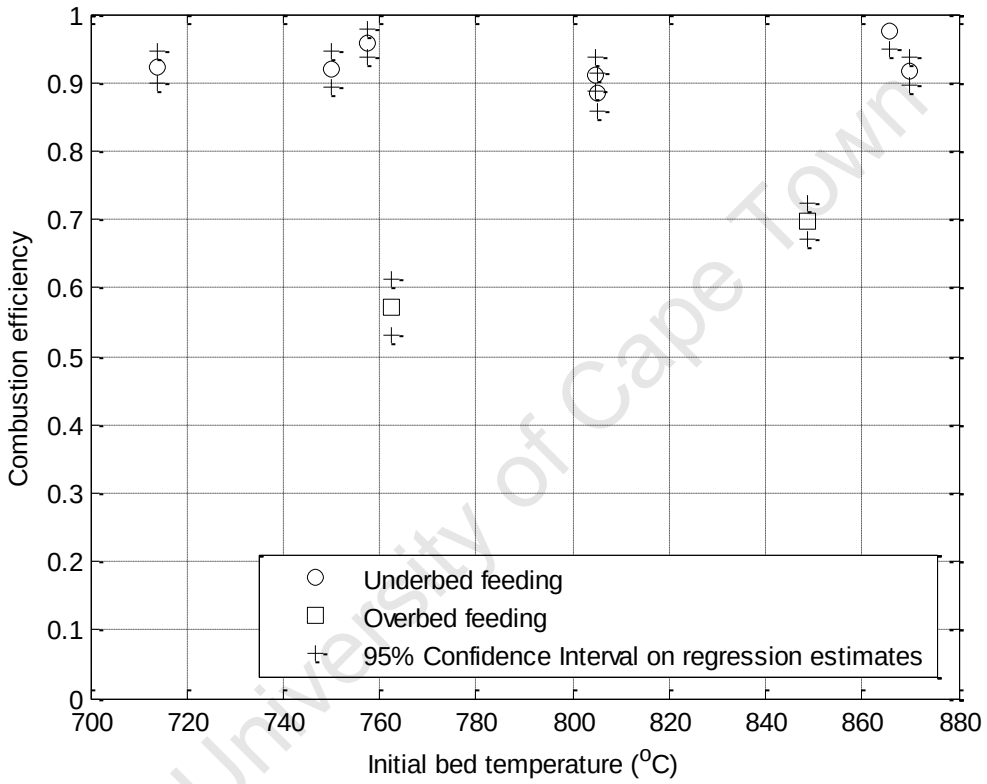


Figure 4-17: Effect of bed temperature on combustion efficiency for wood runs

Figure 4-18 shows the jump in freeboard temperature seen in the underbed and overbed wood runs as feeding commenced. The jump in temperature for the underbed runs is considerably less than the overbed runs (15°C to 30°C compared to approximately 40°C). The overbed runs show little dependence on initial bed temperature. The underbed runs show a marked decrease in freeboard temperature jump as the initial bed temperature increases, dropping from approximately jump of 30°C at a bed temperature of 715°C to 15°C at 870°C.

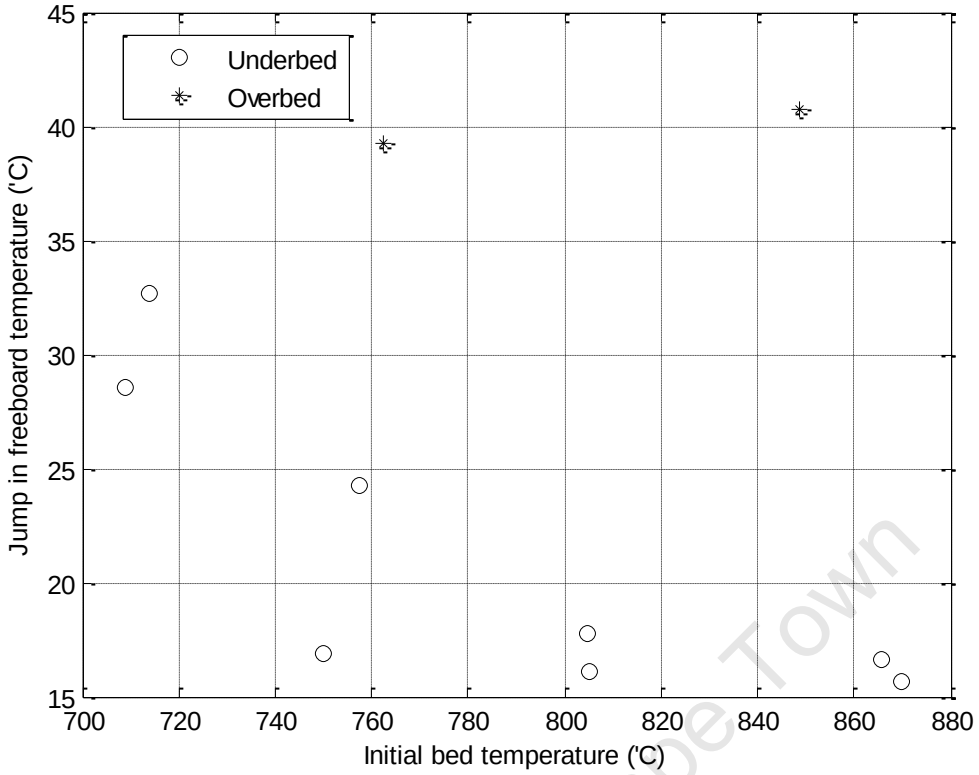
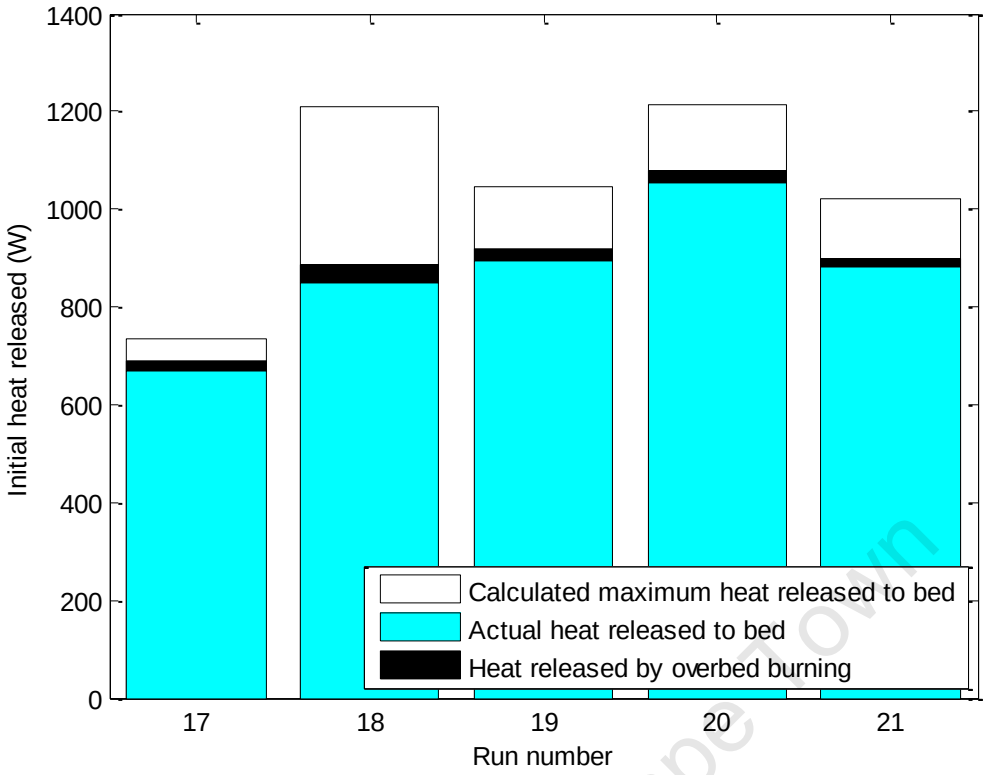


Figure 4-18: Freeboard temperature jump for underbed and overbed feeding of wood

4.2.6.2. Underbed feeding of sludge

Figure 4-19 and Figure 4-20 show the maximum heat release to the bed in absolute and fractional terms, respectively for the under bed feeding of dried sewerage sludge. As seen in Figure 4-19, the maximum heat release to the bed ranges from 730 W to 1200 W, depending on the feed rate. The experimentally observed heat release to the bed ranges from 670 to 1100 W and the average heat release to the freeboard for these underbed sludge runs is 26 W which is less than 10% of the calculated maximum heat heat input. The unaccounted for heat loss in these underbed sludge runs ranges from 43 to 320 W. Figure 4-20 shows a step change from 70% to 90% in combustion efficiency between 715°C and 760°C.

Figure 4-21 shows the jump in freeboard temperature seen in the sludge runs as feeding of fuel commenced. At the an initial bed of 715°C the freeboard temperature is greater than 50°C. It drops to less than 30°C at a bed temperature of 850°C.



Run number	Feed location	Initial bed temperature (°C)	Fuel flowrate (g/min)	U/Umf	ER
17	Underbed feeding	760	2.3	6.2	1.5
18	Underbed feeding	715	3.8	6.7	1.1
19	Underbed feeding	759	3.3	6.9	1.2
20	Underbed feeding	800	3.8	7.0	1.0
21	Underbed feeding	848	3.2	7.2	1.1

Figure 4-19: Partition of heat release between bed and freeboard for sludge runs

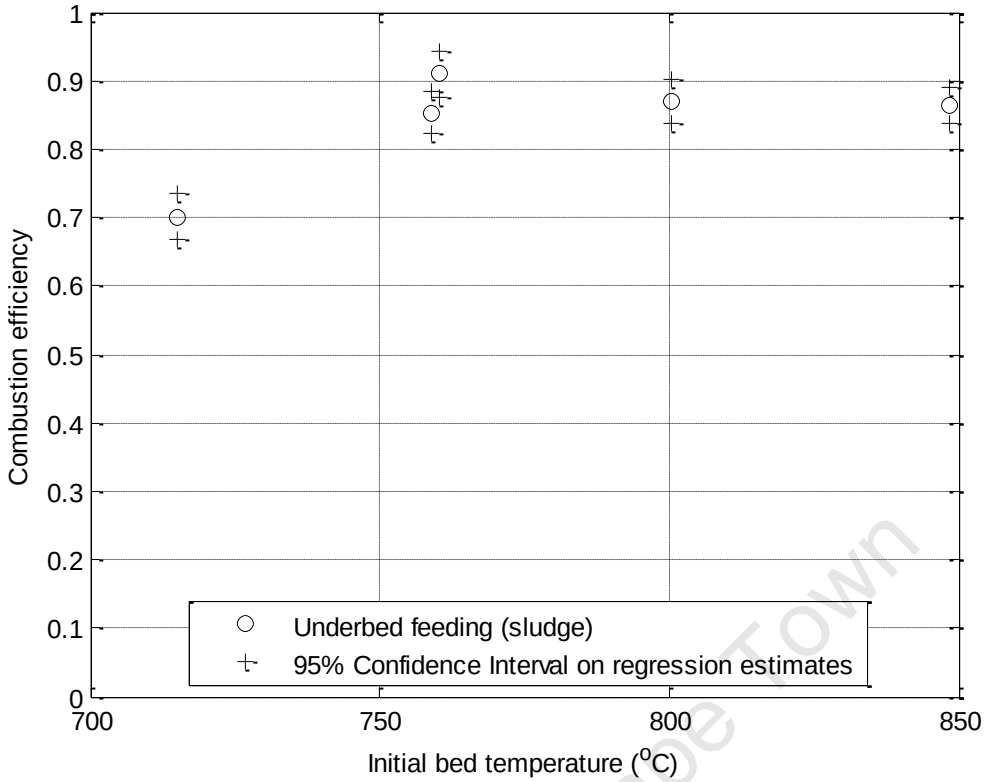


Figure 4-20: Effect of bed temperature on combustion efficiency for underbed feeding of sludge

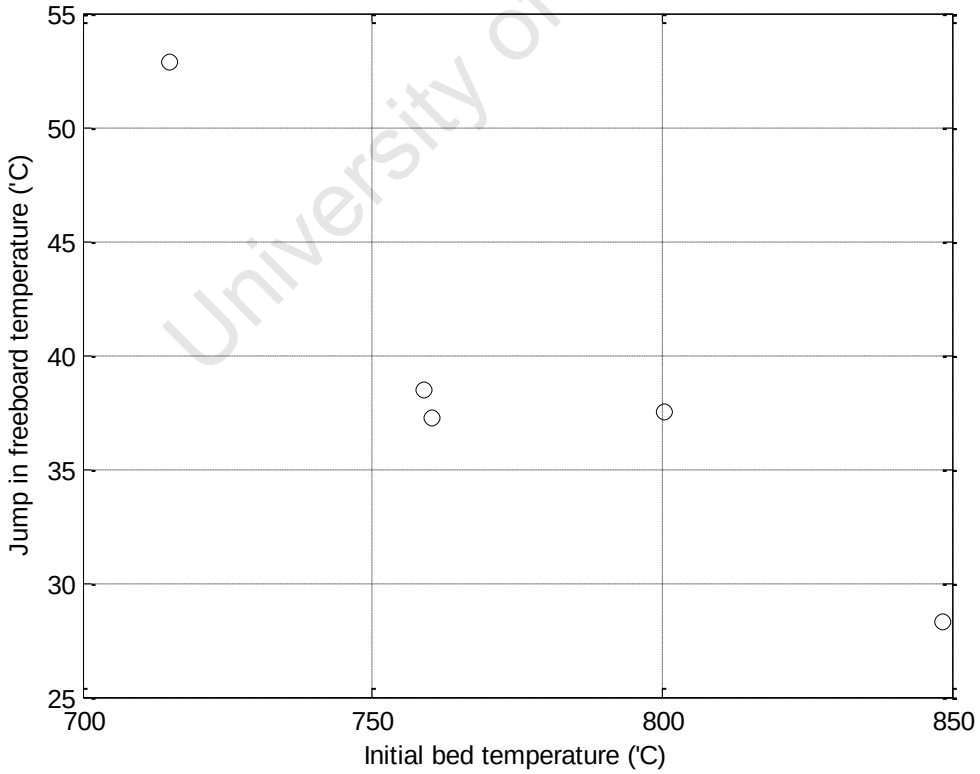


Figure 4-21: Freeboard temperature jump for underbed sludge runs

4.2.7. Discussion

For all runs it is seen that the contribution of overbed burning to overall heat release is less than 10%. This is due to the small thermal mass of the air flowing through the freeboard. However the temperature jump in freeboard was significant, particularly for the overbed wood runs (Figure 4-18). This shows that only a small amount of overbed burning will result in a large temperature jump in the freeboard. Overbed burning thus has implications not so much for energy efficiency, but rather for furnace design such as materials of construction and sizing or placement of heating tubes.

The combustion efficiency for the underbed feeding of wood was 90% and there was little dependence on initial bed temperature, as seen in Figure 4-17. This shows that the temperature was high enough for combustion to occur in the bed, *i.e.* $T_{bed} > T_{crit}$. For the overbed wood runs, presented in Figure 4-17, it is difficult to say whether the increase in combustion efficiency at higher temperature is due to experimental variation or not. More data points are required to comment on a possible trend.

For the sewerage sludge runs, presented in Figure 4-20, the step change in combustion efficiency occurred at some temperature between 715°C and 750° which could indicate that critical temperature for inbed combustion for sewerage sludge exists in that temperature range. This result could have significance for the sludge combustor of Thames Water which operates at a bed temperature of 750°C. Increasing the setpoint would allow the combustor to operate further away from T_{crit}, and hence with higher efficiency.

The feed location of the wood influenced the unaccounted for heat losses and combustion efficiency significantly. When changing from underbed to overbed feeding, the combustion efficiency dropped from 90 to 60%, as seen in Figure 4-17. Overbed feeding results in a decreased residence time in the bed as fuel particles are not forced to travel through the bed but may "float" on the top of the bed. Hence in-bed volatile combustion and tar destruction is compromised, resulting in heat losses due to unconverted material. Qualitatively, it was seen that the overbed wood runs produced more char fines and tar than the underbed runs.

4.2.8. Conclusions and recommendations

An energy balance was used to determine the combustion efficiency as a function of initial bed temperature and feed location. The simplicity of the model was due to the fast rate of reaction of the biomass fuel and the fact that the rate of combustion jumped to a steady state on commencement of feeding. The complex heat transfer of the experimental setup was ignored by considering only the perturbation point, *i.e.* the point at which feeding commenced. It is recommended that for future work on this apparatus, the outer reactor wall temperature is monitored, or the furnace temperature controlled, to allow Q_{in} to be directly measured.

This set of experiments has shown that under the conditions of the investigation, significant overbed burning of the fuels did not occur. In all cases the contribution of overbed burning was less than 10%. The effect of the freeboard temperature jump needs to be considered, if not for energy efficiency purposes, then for furnace design,.

In the case of the underbed wood runs, no clear dependence of combustion efficiency on bed temperature emerged, which could mean that T_{crit} for wood is lower than 700°C. For sludge, the combustion efficiency was ~90%.at temperatures greater than 715°C The step change could be due to the presence of T_{crit} for sewerage sludge in that temperature range. This has implications for choosing an operating temperature for a sludge combustor, as more stable operation would be achieved at higher temperature.

Overbed burning of wood resulted in the combustion efficiency dropping from 90% to ~65%. This is attributed to overbed-fed fuel having a shorter residence time in the bed, resulting in reduced volatiles and tar conversion. The advantages of underbed feeding have been recognised at Thames Water, with the re-design of their sludge incinerator including nozzles to allow for under bed feeding of sludge.

Chapter 5 Processing biogenic waste in South Africa: technology selection and feedstock availability

5.1. Introduction

The White paper on renewable energy in South Africa (DME, 2004) identifies biogenic waste streams as one of the sources of renewable energy which should be harnessed to achieve a target of 10 000 GWhr (1141 MW) by 2013. Biogenic waste streams represent a form of renewable energy which does not compete with food crops nor does its utilisation require the clearing of land. The generation of electricity from biogenic waste streams may be achieved through combustion to raise steam or via anaerobic digestion to produce methane-rich biogas. This work sets out to compare the energy yield from each process by considering feedstock characteristics *viz.* lower heating value, solids content and digestibility. The methodology is then applied to a range of biogenic waste streams in South Africa and the energy potential from these streams is estimated.

5.1.1. Biogenic waste availability in South Africa: previous studies

Estimates of the availability of waste biomass resources in South Africa are limited. Four studies are presented in Table 5-1. Williams and Eberhard (1988) estimated the available energy from biogenic waste sources at 233 PJ/yr while Lynd *et al.* (2003) provided a much higher estimate of 1470 PJ/yr, owing to their proposed role for energy crops. Excluding these energy crops, their estimate is reduced to 300 PJ/yr. Using the energy yields presented in the White Paper on Renewable Energy (2004), the amount was estimated as approximately 280 PJ/yr (see Appendix B for details). In a study commissioned by the WRC, Burton *et al.* (2007) analysed the potential for energy recovery from wastewater streams and animal husbandry in South Africa. In this study, a potential energy recovery from industrial and domestic wastewater of 29 PJ/yr was proposed with the majority harnessed from domestic wastewater, as shown in Table 5-2. The potential energy recovery from intensive animal husbandry was calculated to be 76 to 230 PJ/yr.

Table 5-1: Energy availability from waste biomass in South Africa

	Williams and Eberhard, 1988	DME, 2004	Lynd *et al*., 2003	Burton *et al.*, 2007
Residue	PJ/yr	PJ/yr	PJ/yr	PJ/yr
Agricultural residues	39.4	143.5	156	
Forestry residues	5.15		69	
Pulp and paper and sawmills	26.9	43.8	18	
Sugar industry	55.4	47.6	58	
Cane trash and tops				
Animal husbandry	7.6	20.2		76 - 230
Sewerage sludge	1.8			
Incinerating MSW	96.3			
Energy crops		25.2	1,170	
Industrial wastewaters				29
Total (PJ/yr)	***233***	***280***	***1,471***	***105 - 259***
Total (MW)	***7,374***	***8,889***	***46,645***	***3,300 – 8,200***

Table 5-2: Energy available from wastewater streams in South Africa (Burton *et al.*, 2007)

	PJ/yr	MW
Brewing	0.54	17
Dairy	0.04	1.2
Red meat and poultry abattoirs	0.88	28
Pulp and paper - integrated mills and non-integrated	2.29	72.5
Petrochemical synfuel refineries	1.51	48
Wine and spirits	0.09	3
Fruit processing	2.14	68
Domestic wastewater	21.29	675
Total	**29**	**238**

5.1.2. Harnessing energy from biogenic wastes

Mature energy conversion technologies such as combustion and anaerobic digestion have traditionally been designed for fossil fuels and wastewater treatment systems, respectively. Their role as energy conversion technologies for biogenic waste fuels is a relatively recent development. This implies that it is important to understand their limitations and regions of application for energy generation from biogenic feedstocks.

Lignocellulosic biomass combustors are commercially available and widely used in the USA and Scandanavia to generate renewable electricity (Rosillo-Calle, *et al.* , 2000). In the combustion of biogenic sludges, *e.g.* sewerage sludge, the heating value of mechanically

dewatered sludge may not be sufficient for autothermal combustion making an auxiliary fuel essential (Werther and Ogada, 1999). The relative simplicity of the combustion processes are negated by low thermal efficiencies of 15 to 20% in small boilers (Bridgewater, 2003; McKendry, 2002; Rosillo-Calle, 2000), 30% for larger boilers (Bridgewater, 2003) and 40% for co-firing in coal power plants (Mc Kendry, 2002). Combustion is non-selective in its use of the feedstock, and reduces the whole fuel to simple products (Jenkins *et al.*, 1998).

Reciprocating engines for CHP production from biogas are commercially available. The capacity of these units ranges from 5 to 50 000 kW (Chambers and Potter, 2002). The rate limiting step in the production of biogas is hydrolysis of polymers such as polysaccharides, lipids and proteins (Noike *et al.*, 1985). Anaerobic digestion of wastewater streams usually requires hydrolysis of relatively simple sugars and readily digestible material to reduce COD. However, in energy from waste applications, more recalcitrant feedstocks are processed *e.g* .sewerage sludge and manure slurries.

5.1.3. Objectives

In this chapter, a simple methodology is sought, founded on first principles, to guide determination of the operating region in which combustion and anaerobic digestion can be utilised for the generation of electricity from biogenic waste materials. The methodology developed is then used determine whether combustion or anaerobic digestion should be used to process a variety of biogenic waste streams in South Africa.

The variation in the estimates presented in previous studies highlight the need to further investigate some of the underlying assumptions on which these estimates are based; particularly the larger estimates (e.g. the biofuel contribution in Lynd *et al.* (2003) and the animal husbandry estimates in Burton *et al.* (2007)). The objective of this review is thus to quantify biogenic waste streams in South Africa and calculate the potential energy output. Furthermore, critical discussion of the availability of the stream for energetic purposes is included. Where the stream is already used for energetic purposes, comments on the efficiency are made. Where alternate uses do not preclude energy generation from the resource, its potential is established.

5.2. Methodology to compare combustion and anaerobic digestion energy yield from biogenic wastes

5.2.1. Development of methodology

The dry solids content of the feedstock determines if a positive energy balance is possible over a combustion unit and hence is the first parameter to be considered when comparing technologies. The energy yield from combustion on a dry solids basis is given by Equation 5-1. As stated in Section 3.4.1, the energy yield from the combustion processes is calculated as the difference between the heat released through combustion of solids, as given by the lower heating value of the fuel, and the heat required for evaporating the moisture. Equation 5-2 can be used where it is necessary to account for the ash content of the fuel explicitly.

$$E_{combustion} = m_{DS} \cdot \left(LHV_{fuel} - \frac{(1-DS)}{DS} \cdot \Delta Hvap_{water} \right) \qquad \textbf{Equation 5-1}$$

where $E_{combustion}$ = energy yield from combustion [MJ/day]
m_{DS} = mass flowrate of dry solids [kg/day]
DS = dry solids content (%)
LHV_{fuel} = lower heating value of fuel on dry basis [MJ/kg]
$\Delta Hvap_{water}$ = heat of evaporation of water at 25°C [2.38 MJ/kg] (Perry *et al., 1999)*

$$E_{combustion} = m_{DS} \cdot \left(LHV_{fuel} \cdot (1-ash) - \frac{(1-DS)}{DS} \cdot \Delta Hvap_{water} \right) \qquad \textbf{Equation 5-2}$$

where $E_{combustion}$ = energy yield from combustion [MJ/day]
m_{DS} = mass flowrate of dry solids [kg/day]
DS = dry solids content (%)
ash = ash content of fuel (%)
LHV_{fuel} = lower heating value of fuel on dry basis [MJ/kg]
$\Delta Hvap_{water}$ = heat of evaporation of water at 25°C [2.38 MJ/kg] (Perry *et al., 1999)*

The energy yield from anaerobic digestion (Equation 5-3) can be calculated from the product of the extent of conversion of volatile matter in the feedstock, the specific methane yield and the heating value of methane. As in Section 3.4.1, a specific methane yield of 0.65 m^3/kg volatile solids is used. Alternatively, if the COD (Chemical Oxygen Demand) and volumetric flows are known, as is often the case with wastewater streams, Equation 5-4 can be used to calculate the energy yield.

$$E_{AD} = m_{DS} \cdot (1 - ash) \cdot VSd \cdot Y_{CH_4} \cdot LHV_{CH_4}$$ **Equation 5-3**

where E_{AD} = energy yield from anaerobic digestion [MJ/day]
m_{DS} = mass flowrate of dry solids [kg/day]
VSd = percentage volatile solids destruction (%)
ash = ash content in fuel on dry basis (%)
Y_{CH4} = yield of methane STP [0.65 m^3/kg VS]
LHV_{CH4} = lower heating value of CH_4 at STP [35.8 MJ/m^3] (Perry *et al., 1999)*

$$E_{AD} = Q \cdot [COD] \cdot Y_{CH_4} \cdot LHV_{CH_4} \text{ dfd}$$ **Equation 5-4**

where E_{AD} = energy yield from anaerobic digestion [W]
Q = volumetric flow of stream [m^3/s]
[COD] = COD load of stream [kg COD/m^3]
Y_{CH4} = methane yield at STP [0.35 m^3/kg COD]
LHV_{CH4} = lower heating value of methane at STP [35.8 MJ/m^3] (Tchobanoglous *et al.*,2003)

As seen in Figure 5-1, energy yield from combustion is a non-linear function of dry solids content and thus positive energy yields (heat generated) and negative energy yields (heat required) can be defined. In Figure 5-1, positive and negative energy yields are expressed as a function of dry solids content and LHV for various ash contents. The region above the curve is the feasible region of operation (i.e. positive energy yield) and region below the curve is an infeasible region of operation (negative energy yield). It is clear that increasing ash content reduces the feasible region of operation for combustors. Displayed in Figure 5-1 is a maximum dry solids content likely to be seen for sludge-type biogenic waste. This is based on best operation of dewatering units such as belt filter presses, filter presses and centrifuges (Bane, 2000). When dealing with lignocellulosic wastes, solids contents may reach 80%. An upper limit on heating value for biogenic waste fuels of 20 MJ/kg (dry, ash-free basis) is suggested by Quaark *et al.* (1999).

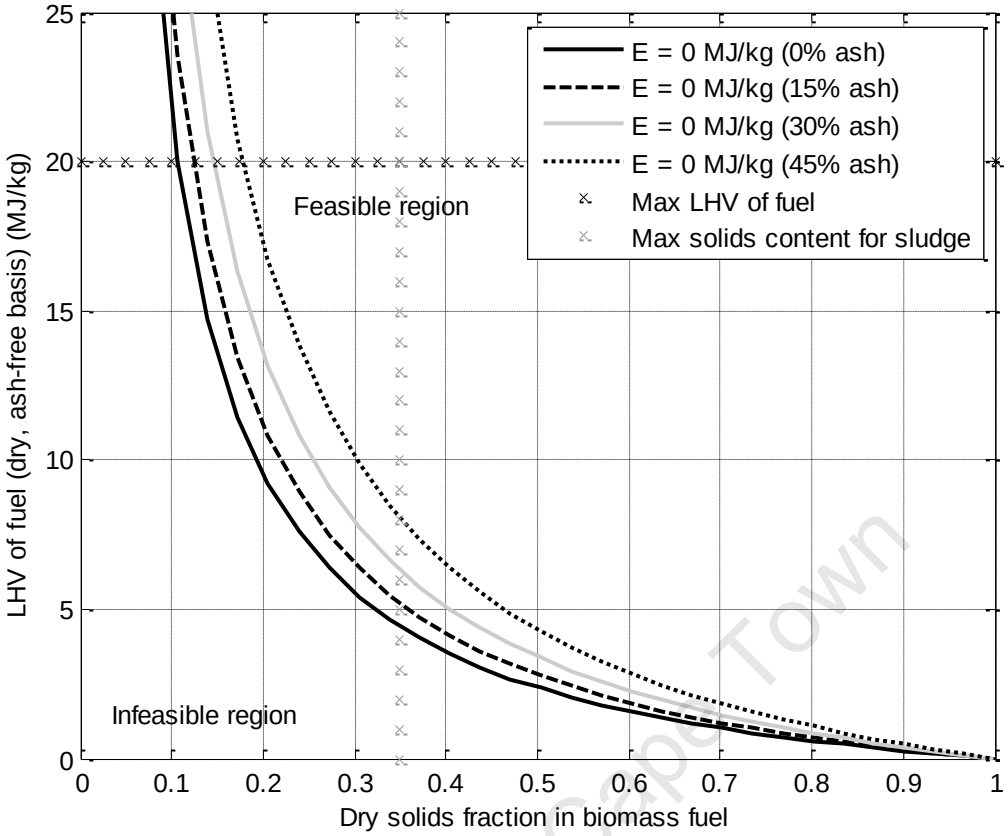


Figure 5-1: Technology selection plot showing feasible and infeasible operating regions for combustion systems

A comparison of the specific energy yield (MJ/dry ton) from combustion and anaerobic digestion can be made using Figure 5-2. The energy yield from combustion is given as a function of dry solids content at different lower heating values. The energy yield from anaerobic digestion is given as a function of volatile solids destruction at different ash contents. The application of Figure 5-2 is shown with three hypothetical fuels: industrial wastewater, biogenic sludge and a lignocellulosic waste. Fuel characteristics are shown in Table 5-3. From these comparisons three points are clear.

- Wastewaters are only energetically suited to anaerobic digestion. Energy yields of 13 to 19 MJ/ton dry solids can be expected depending on volatile solids destruction achieved.
- The energy yield from the combustion of biogenic sludges ranges from 0 to 11 MJ/kg dry solids depending on the dry solids content. The energy yield from the anaerobic digestion of biogenic sludges is 5 to 10 MJ/kg dry solids. Thus, the energy yield of the two conversion processes is similar. The sensitivity of the energy yield from combustion is highlighted.

- Lignocellulosic wastes with their relatively high dry solids content are better suited to combustion, where energy yields of 18 to 20 MJ/kg can be expected. The recalcitrant nature of these materials further promotes them to combustion rather than anaerobic digestion in the absence of pretreatment.

Table 5-3: Fuel characteristics of three hypothetical fuels

	A	B	C
Stream type	Industrial wastewater	Biogenic sludge	Lignocellulosic waste
Volatile solids destruction	60 to 90%	30 to 60%	Low
Solids content	>0.003%	10 to 20%	60 to 85%
Ash content	0%	30%	5%
LHV (dry basis)	N/A	13 MJ/kg	15 MJ/kg

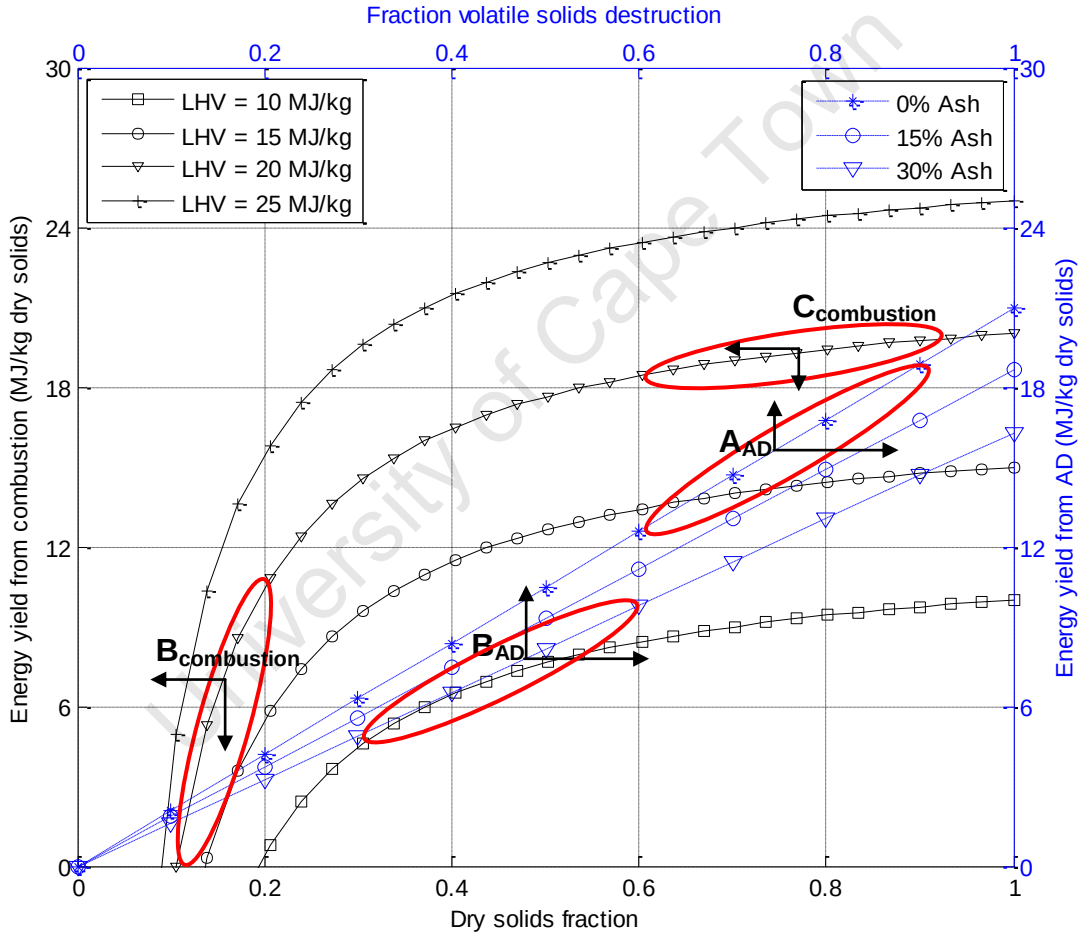


Figure 5-2: Comparison of specific energy yield from anaerobic digestion and combustion for three hypothetical fuels described in Table 5.3

5.2.2. Application to South African biogenic waste streams

Table 5-4 gives the characteristics of a number of biogenic waste streams in South Africa. By superimposing these characteristics on Figure 5-1, it can be seen in Figure 5-3 which streams are suited to combustion.

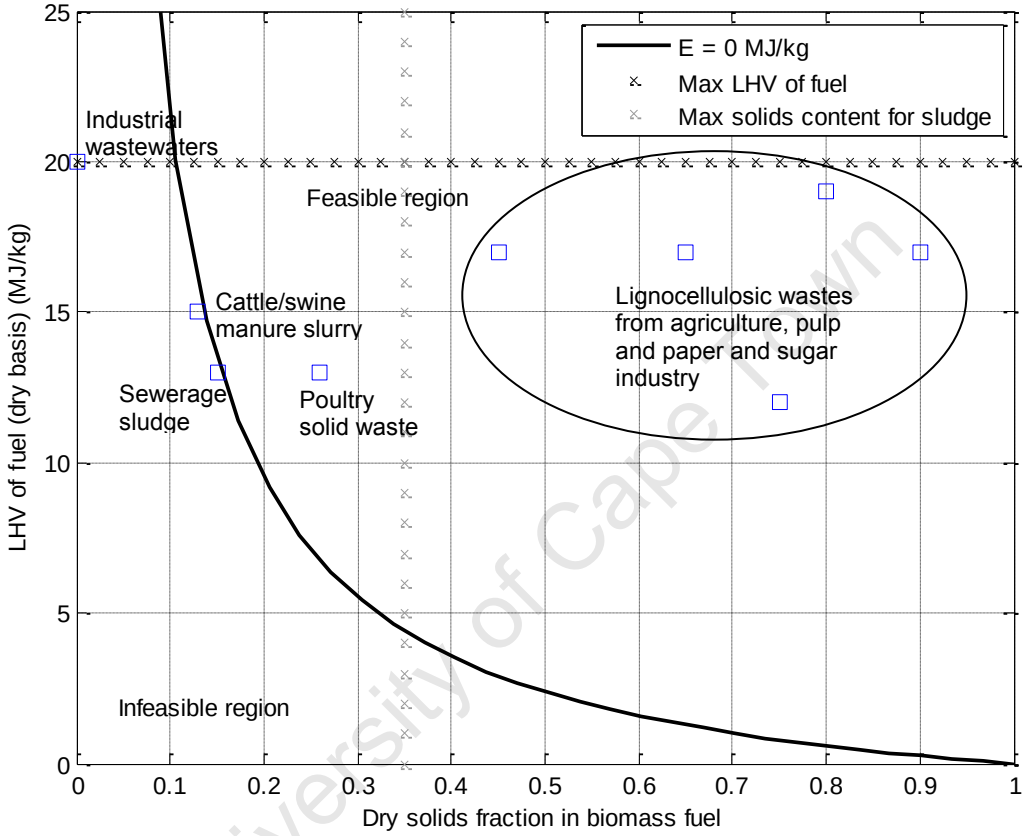


Figure 5-3: South African biogenic waste streams super-imposed on technology selection plot for combustion

Table 5-4: Feedstock characteristics of a number of South African biogenic waste streams

Sector		Municipal wastewater	Industrial wastewaters	Intensive animal husbandry		Agriculture	Paper industry				Sugar industry	
Waste stream		Waste activated sludge	High COD wastewaters	Cattle and swine slurry	Poultry solid waste	Maize and wheat residues	Strong black liquor	Bark	Forestry residues, partly dried in forest	Sawmills	Bagasse	Cane trash and tops, partly dried in field
Energy conversion process		AD	AD	AD	Combustion	Combustion	Combustion	Combustion	Combustion	Combustion	Combustion	Combustion
Load	kg COD/m^3		3.0									
Solids content	%	15%	< 0.03%	13% [a]	26% [a]	90%	75%	80%	65% [b]	65% [b]	45% [c]	70%
Ash content	%	30%	0%	20%	25%	7%	30%	4%	5%	4%	4% [c]	4%
LHV (dry basis)	MJ/kg	13	NA	15	13	17	12	19	17	17	17	17 [d]

a.) Schafer and Wallis, 2005
b.) Foelkel, 2007
c.) Magasiner and de Kock, 1987
d.) Beeharry, 2001
e.) Rosillo-Calle *et al.*, 2001
Unless otherwise stated, LHV's, dry solids contents and ash contents are taken from ECN, 2008.

5.3. Quantifying biogenic waste streams in South Africa

5.3.1. Domestic wastewater

5.3.1.1. Industry overview

The wastewater treatment process has been covered in detail in Section 3.2. In South Africa, the municipalities are responsible for the operation of wastewater treatment plants and the provision of clean water. This is in contrast with the UK where some of the largest water providers are private companies *e.g.* Thames Water.

From the study presented in Chapter 3, it is clear that a significant amount of energy is recoverable from the sewerage sludge generated on a domestic wastewater treatment plant, either via incineration or anaerobic digestion. In the wastewater treatment process, it is possible to use anaerobic digestion to treat the influent stream, as opposed to aerobic processes such as the activated sludge process or trickle bed filters. Indeed, anaerobic wastewater treatment has a number of advantages over aerobic treatment, viz. less energy required; less biological sludge produced and production of a useful product, methane (Tchobanoglous *et al.*, 2003; Waites et al, 2001). However, for domestic wastewaters with lower concentrations of biodegradable COD, higher effluent quality needs and nutrient removal requirements, aerobic processes are favoured as present (Tchobanoglous *et al.*, 2003) and hence, energy recovery from domestic wastewater is not considered in this study.

Approximately 57% of the sludge generated in South Africa WWTP's is anaerobically digested (Snyman *et al.*, 2004). In South Africa the use of the biogas varies between municipalities. Johannesburg and Cape Town do not monitor biogas production and hence it is unlikely they use it. The Cape Flats wastewater treatment plant in Cape Town, which has a granulating operation for the anaerobically digested sludge, was set up to use the biogas in the drying process. Currently this seldom happens and diesel is used instead (Burton *et al.*, 2008). The Tshwane works use the biogas for raising steam to heat the digesters (Ntsowe, 2008). South Africa has one installed sludge incinerator at the eThekwini sewerage works. This unit is currently not operational (Dildar, 2008).

5.3.1.2. Residue production

The total domestic wastewater treated is estimated in Burton *et al.* (2007) as 7600 Ml/yr. In their comprehensive field work on metals in South African sewerage sludges, Snyman *et al.* (2004) drew up a table of wastewater influent volumes and mass of sludge produced from some 70 plants. If sludge production was not measured at the plant, an estimate was based on per capita sludge production. From this data sludge production was calculated to be 282 kg solids/Ml influent volume. The total sludge production for South Africa was calculated to be 782 000 dry ton/yr.

5.3.2. Industrial wastewaters

Data on the COD load and effluent volumes was collected from a series of projects called National Survey (Natsurv) (1989 to 2005), carried out by the Water Research Council. The purpose of Natsurv was to determine current specific water usages and suggest target specific water usages. As such, effluent flows and pollution loads (COD loads) were available. These are listed in Table 5-5. Data prior to 1993 were considered outdated and scaled up as shown. The total quantity of industrial wastewaters is estimated to be 127 Mm^3/yr at an average COD load of 3 kg COD/m^3.

5.3.3. Intensive animal husbandry

Intensive animal husbandry refers to the rearing of animals in feedlots. Four classes of animal husbandry are considered here: cattle, pigs and poultry (layers and broilers). The quantity of waste generated from animal husbandry is calculated as the product of number of heads, specific manure production and dry solids content of the manure as shown in Table 5-6. The number of roaming beef cattle is approximately 12 million (30 times more than those kept in feedlots), however the energy potential from this manure is not considered due to the difficulty in collecting it and the fact that in rural areas it is mostly used for floors, fuel or fertiliser.

Table 5-5: Energy available from industrial wastewaters

	COD concentration	***Effluent flow***	
	mg COD/l	Mm^3/yr	
Brewing	2960 [a.]	8.3	Scaled up according to volumetric flow. Current data from Cole, 2007 cited in Burton *et al.*, 2007
Soft drink	2234 [b]	3.8	Assume same percentage increase in soft drink production as in beer production and hence same increase in effluent production
Dairy	2757 [c]	9.3	Effluent scaled up according to current raw milk production figures as quoted by MPOSA
Sorghum malting	3000 [d]	0.5	Not changed
Sorghum brewing	4000 [d]	1.4	Not changed
Edible oils	4580 [e]	0.6	Not changed
Red meat abattoirs	5000 [f]	5.8	Scaled up based on current number of heads of cattle slaughtered. Current data from Burton *et al.* (2007)
Poultry abattoirs	1760 [g]	7.4	Scaled up based on current number of heads of cattle slaughtered. Current data from Burton *et al.* (2007)
Sugar	1750 [h]	0.4	Not changed as current production similar, as quoted by SASA
Pulp and paper - integrated mills	1430 [i]	60.7	Scaled up according to volumetric flow. Current data from various sources cited in Burton *et al.*, 2007
Non-integrated mills	4550 [i]	10.3	
Petrochemical syn-fuels	384 [j]	3.0	Not changed - data is current
Petrochemical crude oils	384 [j]	1.3	Not changed - data is current
Wine	4714 [k]	0.02	Not changed
Spirits	1714 [k]	0.00	Not changed
Fruit processing	10000 [l]	14 [l]	Not changed - data is current
Weighted average/Total	3046	127	

a. Natsurv 1, 1986
b. Natsurv 3, 1987
c. Natsurv 4, 1989
d. Natsurv 5, 1989
e. Natsurv 6, 1989
f. Natsurv 7, 1989
g. Natsurv 9, 1989
h. Natsurv 11, 1990
i. Natsurv 12, 1990
j. Natsurv 15, 2005
k. Natsurv 14, 1993
l. Burton et al, 2007

Table 5-6: Energy recovery from animal husbandry

Animal	Dry solids content [b]	Wet manure production [c]	Heads [d]	Dry mass generated
	%	ton/animal/yr	#	dry ton/yr
Dairy cow	14%	9	620,000	776,000
Pig	10%	1.9	1,650,000	323,000
Layer	25%	0.047	14,000,000	164,000
Broiler	26%	0.024	545,000,000	3,348,000
Beef cattle in feedlots	15%	8.3	420,000	512,000
Total				5,123,000

b.) ECN, 2008
c.) dpi.qld.gov.au/environment
d.) Same data used as in Burton *et al.*, 2007
e.) 8% collection factor
f.) 40% volatile solids destruction

5.3.4. Agricultural residues

In Table 5-11, the data on seed harvest (taken as the average over the past 4 years), harvest indices and harvest loss, collected from the Crop Estimates Committee and various experts in the Agricultural Research Council (ARC) are used to estimate available residues. Harvest indices are the ratio between amount of seed harvested and the total biomass yield. Harvest loss is due to the harvester trimming the plant at a convenient height above ground level and not capturing the leaves. This loss is estimated as 20% of total biomass harvested (Liebenberg, 2008). Available residues were calculated according to Equation 5-5 and are shown in Table 5-7. A total of 8,4 Mton/yr residues are produced from maize and wheat agriculture.

$$M_{residue} = \frac{(1-HI)}{HI} \cdot M_{seed} \cdot (1-H_{loss}) \qquad \textbf{Equation 5-5}$$

where $M_{residue}$ = amount of residue available for collection (ton)
HI = harvest index
M_{seed} = seed harvest (ton)
H_{loss} = harvest loss

Table 5-7: Agricultural residue production

	Harvest (1000 ton)	Harvest index	Harvest loss (%)	Residue (1000 ton/yr)
Maize	8,726	0.5	20%	6,981
Wheat	1,808	0.5	20%	1,446
Total (1000 ton/yr)				8,427

5.3.5. Forestry

5.3.5.1. Industry overview

South Africa has approximately 1.3 million Ha of plantations which produce hardwoods (eucalyptus and wattle) and softwoods (pine), as shown in Figure 5-4. The plantations are grown in short (~10 years) or long rotation (~25 years). The pulp and paper market uses 67 vol% of total wood production and the sawlog industry uses 25 vol%. The remainder is used by in mining timber, poles and charcoal production (Forestry Industry Facts, 2008; Chamberlain *et al.*, 2005a).

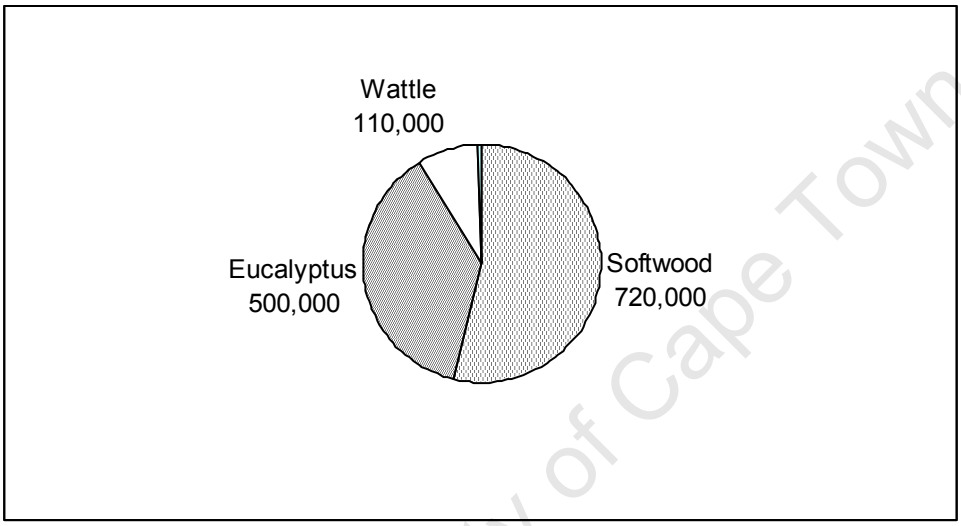


Figure 5-4: Plantation area in Ha for different species (Chamberlain *et al.*,2005a)

5.3.5.2. Residue production

The total yearly sustainable harvest from a plantation can be calculated as the product of the growth rate of the plantation (or Mean Annual Increments (MAI)) and the plantation area. The percentage of the total harvest left in the forest determines the available forest residue. Forest residues comprise wood from pruning, first year trimmings and cuttings from side branches of felled trees. Opinions vary as to the available amount of residue left in the forest. Chamberlain *et al.* (2005b) cite that 10% of total biomass produced is left as residue; Forestry South Africa (FSA) cites 5% as being left in the forest (Edwards, 2008); Lynd *et al.* (2003) assumed 50% in their calculations.

Data from Forestry Industry Facts (2008) and MAI's reported by Chamberlain *et al.*(2005b) are used to calculate the amount of residue produced, as shown in Table 5-8. The following assumptions were made:

- Weighted MAI's were used based on the long and short rotation values given for pine, eucalyptus and wattle (Forestry Industry Facts, 2008)
- Weighted log densities were calculated similarly
- The amount left in the forest was estimated at 10% of total biomass

Using this approach, the amount of forestry residue is estimated at 1.9 Mton/yr. Chamberlain *et al.* (2005b) cite 3 Mton/yr of forestry residues are generated.

Approximately 20% of the forestry residue consists of non-marketable sticks (sticks with a diameter of less than 7.5 cm) which are gathered by surrounding communities for firewood. The leaves, needles and smaller branches left behind are "low heat" burned and the remaining 16% is left as groundcover or mulch (Godsmark, 2005; Theron, 2005 cited in cited in Chamberlain *et al.*, 2005b)

Table 5-8: Energy recovery calculation for forestry residue based on 2005 plantation data

	Plantation area [a]	MAI [b]	Annual harvest	Residue	Log density	Residue
	10^3 Ha	m^3/yr/Ha	10^6 m^3/yr	10^6 m3/yr	ton/m^3	10^6 ton/yr
Softwood	720	14.6	10.5	1.2	1.04	1.2
Eucalyptus	500	21.1	10.6	1.2	0.7	0.8
Wattle	110	12	1.3	0.15	0.8	0.1
Other	7	12	0.084	0.009	0.8	0.007
	1,337		22.5	2.2		1.89

a.) Forestry Industry Facts, 2008
b.) Chamberlain *et al.*, 2005b

5.3.6. Pulp and paper

5.3.6.1. Industry overview

The South African pulp and paper industry is dominated by Sappi and Mondi which hold 62% and 38% of the pulp production capacity, respectively (Chamberlain, 2005a). Sappi has five mills (Saiccor, Stanger, Tugela, Ngodwana and Enstra) and Mondi four (Piet Retief, Durban, Richards Bay and Felixton) (Chamberlain, 2005b). Many of the paper mills generate up to 50% of their energy requirements from renewable sources, *viz.* bark combustors and in the chemical pulping process. Others however, do not feature any energy recovery systems primarily due to the plant age and type of process being used. In these cases, the bark is landfilled and the chemicals are recovered without energy recovery.

Debarking and chipping prepares the logs for the pulping process. Debarking is generally done in a drum debarker and the process efficiency should be in the region of 85-95%. The chipping process is most frequently done by a disk chipper. The chips are screened after production and oversized chips are re-chipped and the sawdust fraction is generally used in the chipboard industry (Ullmans, 2003b).

Pulping extracts the cellulose fibres (pulp) from the matrix of cellulose, hemicellulose and lignin (Ullmans, 2003b). Mechanical pulping techniques expose the cellulose fibres through the mechanical action of grinding stones or refiner disks. Pulp yields of up to 90% are achieved (Chamberlains, 2005b). Chemical pulping techniques use various chemical solutions to dissolve the lignin from the fibre matrix. The raw pulp produced via either method is then washed or bleached before being used for paper-making or other chemical industries. Chemical pulping techniques have lower yields (~50%), however higher quality paper can be produced from chemical pulps. The waste stream contains ligneous substances, dissolved carbohydrates and inorganic pulping chemicals. The stream is concentrated in multiple effect evaporators and then combusted in a modified boiler to recover the chemicals and simultaneously raise steam. Table 5-10 summarises the pulp and paper operations in South Africa.

5.3.6.2. Residue production

By using the yields and pulp production shown in Table 5-10, the total black liquor production can be calculated as 2.8 Mton/yr, of which 95% is currently used in energy recovery boilers. In addition to the black liquor produced, debarking operations also generate a residue stream. Softwoods are generally debarked in the forest and hardwoods are debarked on site. The total amount of bark generated is 0.9 Mton/yr of which 44% is from hardwoods, as seen in Table 5-9.

Table 5-9: Residues from debarking

Type		***Hardwoods***	***Softwoods***
Total log production	ton/yr	8,435,000	10,539,000
Bark content [a]	mass %	6%	6%
Dry solids content [b]	mass %	80%	80%
Bark generated	dry ton/yr	405,000	506,000

a.) Foelkel, 2007
b.) ECN, 2009

Table 5-10: South African pulp and paper mills: pulp production, yields and renewable energy share (adapted from Chamberlains *et al.*, 2005b)

	Process	Plant	Input	Pulp yield	Pulp production (ton/yr)	% renewable energy	Comment
Mondi	Thermo-mechanical	Merebank	Softwood	90%	286,000	53%	Waste stream is organic loaded waste water stream. Anaerobic treatment possible, aerobic used.
	Kraft	Richards Bay	Eucalyptus and pine and small amount waste fibre	50%	575,000	50%	Bark and coal co-combustion practised with plans to increase biomass share from 52,500ton/yr to 125,500 ton/yr in a CDM registered project (DME, 2008)
	Soda	Felixton	Bagasse (80,000); Recovered waste wood fibre (8,000) and waste paper (50,000)	50%	70,000	50%	Further energy recovery unlikely.
	Kraft	Piet Retief	Eucalyptus and pine, purchased pulp and waste fibre (7%) and waste fibre (30%)	50%	60,000	50%	Further energy recovery unlikely.
Sappi	Acid bisulphite	Saiccor	Hardwood timber	35%	600,000 780,000*	39% 45%*	Some lignosulphonates recovered from black liquor. *Recently installed new boiler which reduced imports by 6% even though production increased by 30% (Kerr, 2008)
	Soda	Stanger	Bagasse	50%	60,000	3%	Non-wood mill, hence concentration and concomitant energy recovery from black liquor difficult
	Kraft and NSSC	Tugela	Pine and eucalyptus	50%	350,000	25%	In a CDM registered project 70 000 ton/yr bark (previously landfilled) will replace coal and increase the renewable energy share (DME, 2008)
	Kraft and stone groundwood	Ngodwana	Softwood	50% 85%	510,000	43%	Stoneground wood mill produces high organically loaded wastewater stream. Bark is sent to landfill (Kerr, 2008)
	Soda	Enstra	Eucalyptus timber for hardwood pulping	50%	90,000	2%	Energy recovery possible but historically plant never had and has not installed now

5.3.7. Sawmills

5.3.7.1. Industry overview

The sawmill industry uses predominately long rotation softwood as its fibre input. Mills range from large scale operations to mobile micro mills, the sawn timber volume output and market share are shown in Table 5-11. The industry was regulated by an overarching board called the Timberboard which set prices. This led to lack of innovation and has resulted in South Africa having a sub-standard recovery on the mills *viz.*, 47% compared to 54% (Chamberlain, 2005a). Sawmills produce sawdust and woodchips as waste materials. The large sawmills sell up to 20% of their fibre input as woodchips to the pulp and paper mills (Chamberlain, 2005b).

5.3.7.2. Residue production

The residue production from sawmills is shown in Table 5-11. A total residue of 423,000 ton/yr is generated of which 53% is from large mills.

Table 5-11: Energy recovery opportunities from sawmilling operations

Mill size	***Large***	***Small***	***Micro***
No. of mills [a]	45	106	230
Volume sawlogs produced (m^3) [a]	1,702,000	627,600	174,900
% share of total production	68%	25%	7%
Specific mill production (m^3)	38,000	6,000	800
Recovery [b]	48%	45%	45%
Total log input (m^3/yr)	3,546,000	1,395,000	389,000
Waste generated (m^3/yr)	1,844,000	767,000	214,000
Chip bulk density (dry ton/m^3) [d]	0.2	0.2	0.2
Waste generated (dry ton/yr)	369,000	153,000	43,000
Sold to pulp and paper (ton/yr) [c]	142,000	0	0
Available for energy generation (ton/yr)	227,000	153,000	43,000

a.) Chamberlains et al, 2005a
b.) Based on South African Lumber Mill Index cited in Chamberlains et al, 2005b
c.) 20% of large mill woodchip waste sold to pulp and paper mills (Chamberlains et al, 2005b)
d.) Density of woodchips taken from McKendry, 2002a

5.3.8. Sugar industry

5.3.8.1. Industry overview

Cane is grown primarily in Kwa-Zulu Natal and Mpumalanga. The cane producing areas in KZN are divided into the Coastal Sands and the Northern irrigated areas. Most of the cane in South Africa is produced under rainfed conditions where the annual precipitation is 800 to 1200 mm/yr, making it one of the driest cane producing areas in the world (van Antwerpen, 2006). The national average cane yield is 65 ton/Ha; lower than other cane producing countries such as Mauritius (75 ton/Ha (Beeharry, 1996) and Hawaii (200 ton/Ha (Ullmans, 2003c)).

Cane is harvested mechanically or manually depending on the slope of the land. In South Africa, 90% of cane is harvested dry and prior to harvesting, the trash (dry cane leaves and cane tops) is burnt to make harvesting easier and to reduce the amount of fibre carried into the mill (van Antwerpen *et al.*, 2006). If the cane is harvested green, the trash is left on the field to form a mulch, which returns nutrients to the soil (Singh, 2008). Leaving the trash to dry on the field for a few days can reduce the moisture content to 30%. It can then be raked into windrows, baled and transported to the mill for energy recovery (Rosillo-Calle *et al.*, 2000).

After harvesting the cane is transported to the sugar mill. The cane is initially chopped to pieces 2-10cm by a set of revolving knives and then crushed in a three-roller mill (Ullmans, 2003c). This releases 55-70% of the juice and opens 84-90% of the sucrose-containing parenchyma cells. The cane juice is extracted from the opened cell by lixiviation and from the unopened cells by diffusion. To aid this process, the diffuser temperature is 65-75°C which makes the cells semi-permeable (Baikow, 1982). In the basic operation of a diffuser, bagasse is moved over perforated plates by chains. A lixiviant or "sweet water" is sprayed onto the bagasse and is then collected in a trough under the conveyor (Baikow, 1982). Once the bagasse is exhausted of its sugar, it is dewatered to approximately 50% moisture content before being fed to the bagasse boilers (Ullmans, 2003c). The cane juice (which has solids content of 10-25%) is thickened and then liquid is concentrated to 70% solids in a multi-effect evaporator plant. The resulting cane syrup is then evaporated to supersaturation and seeded to initiate crystallisation. The sugar leaving the mill is known as raw sugar and is sent to sugar refineries which further process the sugar to end-consumer products (Ullmans, 2003c).

There are ten sugar mills in South Africa: two owned by TSB (Malelane and Komati) four by Illovo (Eston, Umzimkulu, Noodsberg and Pongola) and four by Tongaat-Hulett (Maidstone, Amatikulu, Felixton and Darnell). Currently, the Felixton, Maidstone and Amatikulu mills export 9 MW (huletts.co.za, 2008) and the go-ahead for a boiler expansion project to export 38 MW in crushing season and 17 MW out of season depends on Eskom price negotiations (Tongaat-Hulett Annual Report, 2007). The Illovo and TSB operations in South Africa currently do not export electricity. The Illovo Ubombo mill in Swaziland does export electricity to the local grid.

5.3.8.2. Residue production

Data from SASA (2008) for the period 2003 to 2007 gives an average bagasse production of 6 Mton/yr (wet basis), as seen in Table 5-12. On a dry basis this equates to 2.7 Mton/yr.

Cane trash from cane fields which have not been burnt can be used supplement bagasse. Purchase *et al.* (2008) calculate the dry trash yield to be 153 kg/ton wet cane. Using the data in Table 5-12 the residue from cane trash is calculated as 3 Mton/yr.

Table 5-12: South African bagasse production

Season	Cane processed (wet tons)	Bagasse produced (wet tons)
Aug-07	19,724,000	5,954,000
Jul-06	20,279,000	6,144,000
Jun-05	21,052,000	6,247,000
May-04	19,095,000	5,786,000
Apr-03	20,419,000	6,222,000
Average	20,114,000	6,071,000

5.4. Energy recovery from biogenic waste streams

5.4.1. Results

Table 5-13 collates data from Table 5-4, Figure 5-3 and Section 5.3 to summarise the potential energy generation from biogenic waste streams under a number of scenarios. "Maximum potential" assumes 100% of the stream is recovered from energetic purposes, "Probable potential" considers the likely fraction of the stream to be recovered and "Actual" is the amount of stream which is currently being used for energy generation. The energy generated via combustion or anaerobic digestion is calculated according to Equation 5-1 or Equation 5-3 or in the case of the industrial wastewaters via Equation 5-4.

5.4.2. Discussion

5.4.2.1. Domestic wastewater

Two scenarios were prepared: "immediately possible" is the energy recovery from the amount of biogas currently produced and the "potential" is based on treating all the sludge produced by anaerobic digestion and assuming a higher destruction of volatile solids is achieved. For the former scenario, the potential sludge available for energy conversion is the 57% currently treated by anaerobic digestion (other treatments include aerobic and lagooning) (Snyman et al, 2004). The former scenario assumes a volatile solids destruction of 25% based on data obtained from the Johannesburg and Cape Town sewerage works (Section 3.2.2). From this, a final energy recovery estimate from sludge of 3.2 PJ/yr is found. The potential energy recovery is 8.3 PJ/yr. Burton *et al.* (2007) calculate the available energy from sewerage sludge as 16 to 27 PJ/yr (500 to 850 MW). Simplifying the calculation carried out in Table 5-13 to ignore the inorganic component (ash) of the sludge, assume all sludge is anaerobically digested and assume 100% volatile solids destruction, the total energy recovery predicted increases to 550 MW, consistent with the Burton *et al.* study. However, these simplifications appear unrealistic.

Table 5-13: Potential energy generation from biogenic waste streams in South Africa

Sector		Municipal wastewater		Industrial wastewaters	Intensive animal husbandry		Agriculture		Paper industry						Sugar industry			
Waste stream		Waste activated sludge		High COD wastewaters	Cattle and swine slurry	Poultry solid waste	Maize and wheat residues		Strong black liquor	Bark			Forestry residues, partly dried in forest		Sawmills	Bagasse		Cane trash and tops, partly dried in field
Energy conversion process		AD		AD	AD	Combustion	Combustion		Combustion	Combustion			Combustion		Combustion	Combustion		Combustion
Current use or treatment of waste stream		AD or stockpiling		AD or released to municipal wastewater system	AD (lagooning)	Sale as fertiliser or feed	Maize: insignificant amount burnt on field. Usually baled and sold as fodder (R1/kg). Wheat: burnt on field or baled and sold		Heat recovery for paper drying	Left in forest to protect soil or landfilled. Mondi Richard's Bay and Sappi Tugela have registered CDM projects for bark boilers			Often burnt to allow loggers access. Some KZN plantations have instigated a "no-burn" policy. Mondi plans to collect residues for their CDM project		20% of large mill waste sold to paper mills. Mondi plans to collect chips for their CDM project	Process heating requirements. Felixton, Amatikulu and Maidstone generate electricity		Burnt on field
Energy yield scenarios		Maximum potential (100% stream recovery)	Probable potential (57% stream recovery)	Maximum potential (100% stream recovery; 100% VS destruction)	Maximum potential (100% stream recovery)	Maximum potential (100% stream recovery)	Maximum potential (100% stream recovery)	Probable potential (50% stream recovery)	Currently achieved thermal output. Electrical conversion not practised	Maximum potential (residues from on-site and forest debarking recovered)	Probable potential (recovery of on-site debarking)	Currently achieved/in CDM project pipeline No conversion to electricity	Maximum potential (100% stream recovery)	Probable potential (50% stream recovery)	Probable potential (excluding waste sold to mills and waste from small and micro sawmills)	Maximum potential (100% stream recovery for energetic use)	Currently achieved electricity generation. Thermal output higher	Maximum potential (70% recovered for energetic use)
Quantity generated	dry ton/yr	782,000			1,611,000	3,512,000	8,427,000		2,829,000	911,000			1,890,000		423,000	2,732,000		3,077,000
Volume	m^3/yr			127,000,000														
Load	kg COD/m^3			3.0														
Solids content	%	5-10%	5-10%	< 0.03%	13%	26%	90%	90%	75%	80%	80%	80%	65%	65%	65%	45%	45%	70%
Ash content	%	30%	30%	0%	20%	25%	7%	7%	30%	4%	4%	4%	5%	5%	4%	4%	4%	4%
VS destruction	%	65%	25% [a]	100%	40%	NA	Low due to high lignin content		NA	Low			Low		Low	Low		Low
LHV (dry basis)	MJ/kg	13	13	NA	15	13	17	17	12	19	19	19	17	17	17	17	17	17
Stream recovery	%	100%	57%	100%	100%	100%	100%	50%	95%	100%	44%	13% [b]	100%	50%	40%	100%	8%	70% [d]
Thermal energy yield	PJ/yr	8.3	3.2	4.8	12	22	139	69	30	16.5	7.3	2.2	30	15	3	38	3	34
	MW	260	100	150	380	710	4420	2200	950	520	230	70	940	470	80	1220	60	1090
Electrical energy yield assuming 30% thermal efficiency	PJ/yr	2	1	1	4	7	42	21	9	5	2	1	9	4	1	12	0.9	10
	MW	79	30	46	114	213	1325	660	284	157	69	21	283	141	25	366	29	327

a.) King, 2008; Deacon, 2008
b.) Data from CDM Project Design Documents (DME, 2008) show that Mondi Richards Bay and Sappi Tugela burn (or plan to burn) a total of 0.12 Mton/yr of bark. This is 13% of the total available stream
c.) Felixton, Maidstone and Amatikulu exported 9 MW to the grid (huletts.co.za, 2008) from 4.5 Mton cane milled in 2008 (Rawlinson, 2009). On-site requirements are calculated at 10 MW for the 3 mills based on a site-wide usage of 20 kWhr/ton cane processed (Beeharry, 1996; Bhatt and Rajkumar, 2001)
d.) Beeharry, 2001 suggest 30% be left in the fields

5.4.2.2. Industrial wastewater

The total energy available from wastewater streams was calculated according to Equation 5-4 to be 4.8 PJ/yr (150 MW). This is in the same range as the value of 238 MW calculated by Burton *et al.* (2007). The discrepancy between these studies arises partly from the energy estimated as recoverable from synfuel petrochemical refineries. Burton *et al.* (2007) attribute 48 MW to synfuel refineries, based on the PetroSA gas-to-liquids refinery energy yield per barrel of synfuel produced which is being achieved in their CDM project, and scaled up to total South African synfuel production. The calculation of energy recovery from petrochemical refineries in this work is based on an influent COD of approximately 400 mg COD/l (Natsurv 15, 2005). This value is uncharacteristically low for an industrial wastewater stream and is thus the possible source of error. Indeed, work on anaerobic digestion of the Sasol coal-to-liquids refinery wastewater gives a wastewater COD of 11 000 mg COD/l (Britz *et al.*, 1983).

In terms of national consumption, the potential energy from wastewaters is a small contribution, approximately 0.6% of national demand. Further, to date, the primary goal of the industries producing the wastewaters has been to reduce COD loading to avoid municipal fines. This reduction in COD can be achieved via aerobic and mechanical processes, hence energy recovery from wastewater streams has been further reduced. However, the current focus on renewable energy and on the reduction of CO_2 emissions demands renewed interest in AD. In terms of reducing COD, anaerobic digestion does hold one distinct advantage, namely the generation of CER's under the Kyoto Protocol Emission trading scheme. Projects registered as CDM projects include the PetroSA biogas to electricity project and the Rosslyn Brewery fuel switch project (coal to biogas and natural gas) (DME, 2008)

5.4.2.3. Animal husbandry

The energy recovery from animal husbandry activities is calculated to be 12 PJ/yr from anaerobic digestion of cattle and swine slurry and 22 PJ/yr from combustion of poultry litter as shown in Table 5-13. This gives a total of 34 PJ/yr (1090 MW); less than half of the lower estimate of Burton *et al.* (2007) but much higher than William and Eberhard's (1988) estimate of 7.6 PJ/yr. The energy potential of animal husbandry wastes has been realised in some instances with two piggery waste-to-biogas projects, namely the Kanhym and the Humphries farm manure projects, awaiting approval from the CDM executive board (DME, 2008).

5.4.2.4. Agriculture

The maximum energy (100% stream recovery) from agricultural residues is calculated as 139 PJ/yr (4420 MW). This compares well to the amount calculated in Lynd et al (2003) of 156 PJ/yr (4900 MW). A probable stream recovery of 50% gives a energy yield of 69 PJ/yr (2200 MW). This is thought to be more realistic as maize stubble is baled as sold as fodder and wheat residues are still burnt in some areas of the Western Cape.

5.4.2.5. Forestry residues

To calculate the amount of energy recoverable from the forestry residues, it was assumed that the residues have been allowed to dry to 35% moisture content, which is mid-range of that suggested by Foelkel (2007). On this basis maximum stream recovery gives 30 PJ/yr and at 50% stream recovery, 15 PJ/yr. Lynd *et al.* (2003) give an estimate of 69 PJ/yr, however they assume 50% of forestry residues are left in the forest which is unrealistic.

Forestry residues are in some areas burnt prior to tree felling to allow the loggers easy access to the trees and to mitigate a fire hazard when the new trees are planted. Mondi initiated a "no-burn" policy in the KZN coastal sands plantations to prevent soil erosion (Howard, 2008) and they plan to collect up to 35 000 ton/yr of residues as part of their CDM project (DME, 2008).

5.4.2.6. Debarking

Three scenarios for energy recovery from debarking operations are given: 100% stream recovery where bark is recovered from both the on-site operations and the in-forest debarking operations, 44% recovery in which it is assumed that the on-site operations recover the bark and the in-forest operations leave the bark as a mulch and finally the amount of bark which is currently being used in bark boiler, *viz.* 13%. The energy yields from the respective scenarios are 16.5 PJ/yr, 7.3 PJ/yr and 2.2 PJ/yr.

5.4.2.7. Pulp and paper industry

Pulp mills (non-integrated mills) are generally energy self-sufficient and able to export electricity and steam to the paper mills; however the paper mills require large quantities of steam to run their driers. In the States, the pulp mills supply up to 50% of the integrated mill's energy requirement (Jeffries, 1997). From Table 5-10 it is seen that four mills reach this benchmark. This figure increases to six mills, when a greater than 40% renewable energy

share is considered. The remaining three mills do not practise energy recovery and account for the stream recovery of 93%. The energy recovered from the pulp mills is 30 PJ/yr and is in line with previous studies, but it is noted that this is directly used in the paper mills to raise steam for the paper driers.

5.4.2.8. Sawmills

It is assumed that only the waste from the large mills is available for energy production as the specific mill production of the small and micro mills does not warrant the installation of a boiler. This accounts for a stream recovery of 40%. The probable energy recovery from sawmills is thus 3 PJ/yr. In their CDM project, Mondi Richards Bay intends to use 35 000 ton/yr realising some 0.2 PJ.yr energy production from sawmill residues (DME, 2008).

5.4.2.9. Sugar industry

Two scenarios are presented for energy recovery from bagasse: 100% stream recovery which yields 38 PJ/yr (slightly lower than previous studies) and the currently realised energy recovery of 2 PJ/yr (8% of total stream). The recovery of cane trash for energetic use is not currently practised but a maximum potential yield of 34 PJ/yr exists, assuming the cane trash is allowed to partially dry in the field as suggested by Rosillo-Calle *et al.* (2000).

Various studies on the sugar industry seek means of improving energy efficiency of the mills. Reid (2006) suggests reasons as to why South African mills continue to import coal for their thermal needs:

- Increased alternative uses of bagasse (e.g. animal feed production, pulp and paper feedstock and furfural production)
- Increased addition of imbibition water to the diffuser to increase sugar extraction, but requires increased evaporative duty downstream
- The addition of pan movement water to the vacuum crystallisers to maintain control of the growth of crystals. If the cane syrup quality is bad the amount of water added has to increase
- Not all boilers have an economiser fitted to preheat the boiler water; this represents a 9% loss in steam efficiency (Magasiner, 1987)

Reid (2006) shows that it would be more economical for the mills to add less imbibition water (and hence achieve less sugar extraction) and make savings on the amount of coal used. Concerns for the downstream processing of a more concentrated cane syrup solution include:

difficulty in clarification of the syrup and fouling in the evaporator tubes due to low rising film velocity. The study shows the drawbacks of independent optimisation of process variables which had clearly taken place on the mill.

Mohee and Beeharry (1996) conducted an LCA to investigate the effect on cane yield if a fraction of the bagasse was composted. It was found that if 24% of the bagasse was composted with all the filter cake, the cane yield increased by 50%. However, whilst the mill could still provide its steam requirements with the remaining bagasse, there was a shortfall in electricity production. Only if the boiler efficiency and plant steam consumption was optimised could electricity production be met (Beeharry, 1996).

Cane trash represents a large potential residue stream, however, the effect of trashing (*i.e.* leaving the green leaves and tops to form a mulch) on cane yield showed a 9.3 ton cane/Ha/yr increase compared to burning. The increase in yield is attributed to the higher soil organic content (van Antwerpen *et al.*, 2001). The trash yields for the various harvesting techniques are shown in Table 5-14. Cane trashing also reduces the amount of evaporation. Thompson (1965, 1966) (cited in van Antwerpen, 2006) found that the potential moisture conservation from trashing was 90 mm/yr, which would have a positive effect on yield, especially in dry years. It is recommended, to fully consider the effects of replacing fertiliser with the use of cane trash, an LCA be conducted to calculate the displaced environmental burden.

Table 5-14: Cane residue yields for different harvesting techniques (van Antwerpen *et al.*, 2001)

	Yield of residue (ton/Ha/yr)	
Treatment	With fertiliser	Without fertiliser
Cane burnt	0	0
Cane burnt, tops spread	3.2	2.3
Trashed (tops and leaves spread)	20	15

5.4.2.10. Comments on the use of energy crops in South Africa

The contribution of energy crops in the analysis of Lynd *et al.* (2003) is based on a study done by Marrison and Larson (1996). These workers calculated an area available for energy crops by adding up area defined as pasture and "other" by the World Resource Institute, where "other" included uncultivated land, non-pastoral grassland, built on areas, wetlands, wastelands, roads and deserts. They excluded deserts but considered everything else to be available. For South Africa, the pasture and "other" excluding deserts was estimated as 104

MHa available for energy crops. In the analysis of Lynd *et al.* (2003), 10% (10 MHa) is assumed to be utilised for energy crops. The figure of 10 MHa does not contain sufficient local knowledge of SA soil type and current land usage be regarded as an accurate estimate which can be set aside for energy crop production. The estimate is essentially based on amount of marginal land. In a submission to the Department of Environment and Tourism, Laker (2005) considers some factors influencing sustainable soil practice.

- South Africa has generally poor quality unstable soils and low rainfall. Only 13% of the land (14 MHa) is considered arable and only 3% is considered to be high potential land
- Considering that 0.4 Ha of land are required to feed one person, South Africa has already exceeded the optimal carrying capacity of its soil
- Only 35% of land is suitable for dryland farming. There is limited irrigation potential in South Africa due to either lack of water and unsuitable soil types
- Whilst soil ecosystems in SA tend to be robust (*i.e.*, they do not degrade easily), they shown little resilience (*i.e.* they cannot recover easily once degraded) and hence policy should focus on preventing extensive soil degradation

Substantial degradation (up to 25%) of South African soils has been reported. Practices which could intensify this trend (such as planting crops in marginal land) should be approached with care.

The issue of land availability much be followed by the considering what to plant. *Jatropha* has been much vaunted as a high-yield, drought tolerant species. The oils of *Jatropha* could be used in biodiesel production, but equally well, the entire plant could be used in combustion and gasification systems. Due to limited data on its water requirements, the Department of Water and Forestry specified that its large scale cultivation be regarded as a Stream Flow Reduction Activity (SFRA) (Holl *et al.*, 2007). This classification essentially prohibits its large scale planting. The Water Research Commission funded a study into the hydrological effects of *Jatropha* plantations, with the specific aim of determining whether the planting really should be regarded as a SFRA. The study brought a couple of important points to light (Holl *et al.*, 2007):

- Areas of low and variable rainfall, and areas susceptible to anything more than mild frost, should be avoided (*i.e. Jatropha* should not be planted in marginal areas). The former is illustrated through Figure 5-5.

- It would appear that *Jatropha* is unlikely to have a negative impact on stream flow and cannot be declared an SFRA on the basis of this research
- *Jatropha* has been presented as a wonder-plant when, in reality, very little is known about it and actual large-scale success stories cannot be found

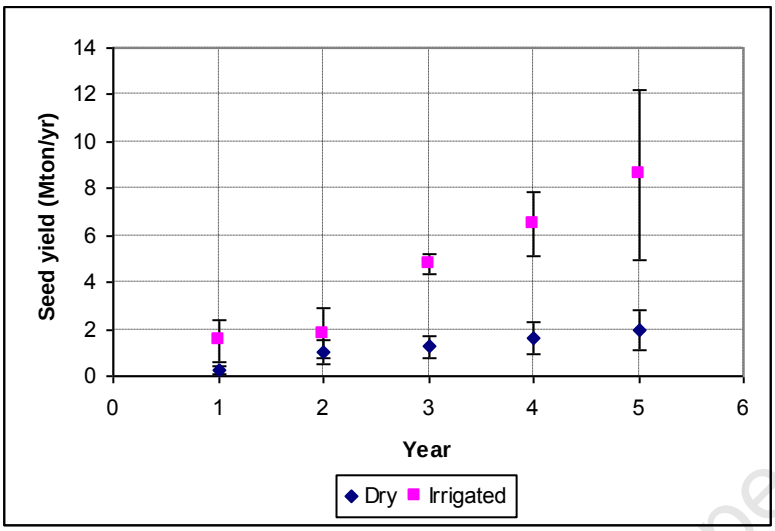


Figure 5-5: *Jatropha* seed yields from dry land and irrigated lands (data from jatrophabiodiesel.org)

High diversity-lowinput grassland may be a more suitable energy crop in South Africa. Tilman *et al.* (2006) compared monocultures of grass to plots with increasing number of species and found a logarithmic relation between number of species planted and biomass energy yield. The experiment was carried out on agriculturally degraded land; additional fertiliser was not added and plots were only irrigated during establishment. In a plot with 16 different species the biomass energy yield was 68 GJ/Ha/yr. This is considerably lower than the DME's figure of 84 GJ/Ha/yr, although the latter is based on high rainfall areas. The White paper considers yields from grassland in the savannah regions to be negligible and thus their approximation of 25 PJ/yr from energy crops is thought to be fair.

5.4.2.11. Comments on using energy recovery from Municipal Solid Waste

The DME (2002) cites energy recovery from Municpal Solid Waste (MSW) as a renewable energy source, but does not give an estimate on the potential energy recovery. Williams and Eberhard (1988) approximate a recovery of 96 PJ/yr by incinerating MSW and von Blottnitz *et al.* (2006) cite 71 PJ/yr based on 9 Mton/yr of MSW generation from six metropoles and an average LHV of 8 MJ/kg. However, municipal solid waste incineration is not practised in South Africa currently, as the waste is sent to landfill.

The anaerobic conditions of the landfill allow for the evolution of landfill gas containing 40% to 60% methane. CDM projects to capture and destroy landfill gas or to capture and use it to generate electricity were seen as the "low hanging fruit" of CDM projects (Strachan *et al.*, 2008). A number exist in the CDM project pipeline and are summarised in Table 5-15. Some of the projects intend to generate electricity to sell back to the grid. For projects not intending to generate electricity, a potential electricity production has been calculated (see Appendix B.3). Currently, 5 MW are being generated (Strachan *et al.*, 2008) and a potential generation of 45 MW exists if all the CH_4 were converted to electricity. It must be noted; however, that the amount of carbon reduction achieved per kg CH_4 destroyed is 21 kgCO_2eq for flaring and only an additional 3.4 kgCO_2eq for generating electricity. It is thus unlikely that many of the flaring projects will move to generating electricity; unless electricity prices increase or the premium for renewable energy increases.

Table 5-15: Electricity generation from landfill sites

Landfill	***City***	***Project status***	***Electricity or flare***	***Carbon reductions in first 7 yrs***	***Potential electricity***
				tCO2 eq	MW
Marrian Hill/La Mercy	eThekwini	Commissioned in Dec 2006	1 MW	480,000	1.2
Bisasar Street	eThekwini	Commissioned in Jan 2008	4-6 MW	2,466,957	6.0
Bellville South	Cape Town	DNA approval pending (PDD submitted)	Planned electricity	129,000	0.3
Chloorkop	Kempton Park	Commissioned Jan 2008	Flare only	1,300,000	3.7
Ekurhuleni	Ekurhuleni	DNA approval pending (PDD submitted)	Flare only	1,700,000	4.8
Tshwane	Tshwane	Project idea approved (PIN submitted)	Planned electricity	5,621,000	13.6
Bulbul drive		Project idea approved (PIN submitted)	Flare only	5,264,000	14.8
Alton	Richards Bay	DNA approval pending (PDD submitted)	0.4 MW	264,430	0.4
Total			***5 MW***		***45 MW***

5.4.3. Conclusions

A methodology to compare the energy yield from combustion and anaerobic digestion based on feedstock characteristics is presented. Following selection of appropriate technology, energy yield equations were used to determine the energy recovery. Competing uses for the residue stream have been highlighted in the study, *e.g.* nutrient recycling in the case of agricultural residues, cane trash and forestry residues. Determining the optimum amount of residue to be left on the land for these purposes is beyond the scope of this project. Obviously, the current practice of burning these residues serves no purpose energetically or otherwise. Pulp and sugar mills currently use their residues for energy recovery. Pulp mills provide, in most cases, 50% of the integrated mill requirement, in line with US mills. Sugar mills generate steam in bagasse boilers for plant-wide consumption. Only on three mills is electricity exported to the grid. Table 5-16 summarises the three scenarios presented in the study: maximum potential, probable potential and currently achieved. Where only one scenario is considered, that value is taken.

Maximum and probable energy recoveries, *viz.* 310 and 200 PJ/yr are similar to those obtained by Williams and Eberhard (1988) and the DME (2004), and to Lynd *et al*. (2003) when the contribution from energy crops is neglected. Assuming a thermal efficiency of 30%, the maximum and probable electrical output from biogenic waste represents 11% and 7.5% of current demand, respectively. If the currently achieved energy recovery were converted to electricity, it would represent 1.3% of current demand. It is thus concluded sourcing 7.5% of current electricity demand from renewable sources is a realistic target.

Table 5-16: Summary of energy recovery potential from SA biogenic waste streams

	Thermal output		Electric output (at 30% efficiency)	
	PJ/yr	MW	PJ/yr	MW
Maximum potential	310	9,800	93	2,900
Probable potential	200	6,300	60	1,900
Currently achieved	34	1,100	10	330

Chapter 6 Conclusions

6.1. Technologies suitable for heat and electricity generation from biogenic waste fuels

South Africa currently generates ~95% of its electricity from fossil and nuclear sources and the remainder from large scale hydropower. The White Paper of the DME in South Africa on renewable energy published in 2003 seeks to increase the share of renewable energy to 4% of the projected 2013 demand. Electricity from biogenic waste sources is identified as contributing to renewable energy supply. Technologies suitable for processing biogenic waste sources to process heat and electricity include combustion, gasification and anaerobic digestion.

Combustion of biogenic waste sources, particularly woody biomass and sewerage sludge, is a well established technology. The process is relatively simple but low thermal efficiencies of 15 to 20% are found in small biomass boilers. This may increase to 30% for larger boilers or to 40% for co-firing in coal power plants. The advantages of fluidised bed combustors for processing waste biomass fuels are highlighted. In particular, uniform bed temperatures are easily achieved due to intimate mixing of the bed material and the large thermal mass of the bed material which maintains temperature and promotes complete carbon conversion. Gasification offers higher efficiencies, but only at higher capacity (100 MW achieves 45% in an IGCC system) and greater system complexity. Indeed, successful installation of large scale biomass gasifiers (greater than 100 MW) is yet to be realised. Technical barriers which exist include difficulties in tar removal from the fuel gas and agglomeration of the bed in fluidised bed gasifiers. Biogas production for energy generation has come to the fore recently although anaerobic digesters are still primarily installed to treat wastewater. Thus, control philosophies may not be set up to optimise continuous biogas production. Nonetheless, the technology is well-established in South Africa, particularly as domestic wastewater treatment plants, in the treatment of synfuel petrochemical waste streams and at the breweries. For the treatment of recalcitrant substances, such as lignocellulosic materials and microbial sludges, pretreatments are available to change the structure of the biomass and make it more amenable to digestion. Of these three technologies reviewed, combustion and anaerobic digestion stand out as

immediately implementable options for South Africa due to the relative simplicity of the process and the operational experience which exists in the country.

6.2. Comparison of combustion and anaerobic digestion for processing biogenic waste – a case study

A comparative case study of three wastewater treatment plants, each employing different technologies to recover energy from waste sewerage sludge was carried out. The technologies employed were incineration (combustion), anaerobic digestion and enhanced anaerobic digestion (*i.e.* anaerobic digestion preceded by a pretreatment, in this case thermal hydrolysis). Field notes on the operation of these plants were presented to expose the South Africa audience to the energy generation potential of wastewater treatment plants and associated difficulties. Energy yield equations for combustion and anaerobic digestion were used to predict the relative gross electricity output and this was compared to plant data. Anaerobic digestion with thermal hydrolysis pretreatment gave the highest energy yield. This was due to the anaerobic digester process achieving a volatile solids destruction of 65% as opposed to 45% as a result of the feed pretreatment. Combustion of a sludge containing 20% dry solids and anaerobic digestion of sludge to achieve 45% volatile solids destruction resulted in comparable energy recovery. However, the plant data showed that combustion was the inferior option. This was attributed to the sensitivity of gross energy recovery to dry solids content, *i.e.* any deviation from 20% dry solids content would have a significant impact on energy recovery. For all technologies, the process heat requirements were supplied by the high grade heat generation. Combustion and anaerobic digestion generated up to five times and double the high grade sensible heat required respectively. This brings into question possible improvements in heat use, e.g. drying the sludge prior to combustion.

6.3. Biogenic waste feedstock characteristics and their effect on energy recovery

The effect of characteristics specific to biogenic waste fuels on their processing was explored. In the case of anaerobic digestion, achieving the maximum biogas yield is prevented by the presence of recalcitrant compounds e.g. cellulose, hemicellulose, lignin and microbical cells. The energy requirements of high temperature, high pressure (HTHP) pretreatments as a function of temperature and dry solids content were measured against the energy yield from anaerobic digestion. This showed the process conditions necessary for the overall process to

be energy positive. The energy requirements of high pressure homogenisation were seen to be one order of magnitude greater than the energy yield from anaerobic digestion at operating pressures cited in literature (30 to 60 MPa). HTHP pretreatments are recommended as the analysis showed a positive nett energy yield at dry solids contents greater than 20% and at volatile solids destruction of greater than 50%, assuming no heat recovery and a thermal efficency of 40%. Heat recovery from the heated pretreated stream reduces the energy requirements and hence makes the pretreatments viable at lower volatile solids destruction.

The high volatile content of biogenic waste fuels negatively affects combustion efficiency in a bubbling fluidised bed through increasing the tendency for overbed burning. Higher bed temperatures can be used to promote inbed combustion. This is desirable as overbed burning results in unstable operation, viz. runaway freeboard temperatures and difficulties in maintaining the bed temperature. The feed location also impacts the combustion efficiency, with under bed feeding promoting a longer residence time of the fuel in the bed. In the experimental study, wood chips and dried sewerage sludge were fed to a bubbling fluidised bed combustor at different temperatures and from different feed locations. No effect of initial bed temperature on combustion efficiency for woodchip was seen. For sewerage sludge, a step change in combustion efficiency was seen between 715°C and 750°C, indicating a change from inbed combustion to overbed burning or the escape of unburnt volatiles and tars. When feeding woodchip overbed, the combustion efficiency dropped dramatically, from 90% to ~65%. It is proposed this was due to higher amount of unburnt volatiles and tar passing through the system unburnt. Notably, this study showed that the contribution to loss in combustion efficiency from overbed burning was small (less than 10%). While the absolute contribution of overbed burning to total energy release is small, freeboard temperature increases of up to 50°C were seen. These two phenomena are reconciled by the relatively small thermal mass of air (compared to the thermal mass of the bed) passing through the freeboard.

6.4. Energy from biogenic waste in South Africa

A methodology is presented to select the technology which gives the highest energy yield based on intrinsic feedstock characteristics. The first tool delineates regions of feasible and infeasible operation for combustion of biogenic feedstocks based on ash content, dry solids content and heating value. The second tool allows comparison of the specific energy output

from combustion and anaerobic digestion based on dry solids content, heating value, volatile solids destruction and ash content.

Following selection of appropriate technology, energy yield models were used to determine the energy recovery. The quantity of residue available was calculated for each waste stream. Competing uses for the residue stream have been highlighted in the study, *e.g.* nutrient recycling in the case of agricultural residues, cane trash and forestry residues. Three scenarios for energy recovery were presented: maximum, probable and current. The maximum and probable energy recoveries calculated in this work, *viz.* 310 and 200 PJ/yr and are similar to those obtained by Williams and Eberhard (1988) and the DME (2004), and to Lynd *et al*. (2003) when the contribution from energy crops is neglected. Assuming a thermal efficiency of 30%, the electrical output from the maximum and probable scenarios from biogenic waste represent 11% and 7.5% of current demand, respectively. The thermal output of the current scenario is 34 PJ/yr (1100 MW). It is thus concluded that sourcing 7.5% of current electricity demand from renewable sources is a realistic target.

Finally, the incremental improvements in the whiteness of our sugar, the fineness of our paper and other perceived product qualities are at the energetic expense of the upstream processes of our industries. The solution may lie not in continuously improved energy efficiency, but in reduced consumer demand for “perfection”.

Reference list

1. Abrams, LJ, 1996, "Water policy development in South Africa", presented at the Cranfield International Water Policy Conference, Cranfield University, Bedford UK, September 1996, accessed from www.africanwater.org, October 2008
2. ACEEE, 2007, abstract from American Council for an Energy Efficient Economy, accessed from www.aceee.org, October 2008
3. afdc.energy.gov, 2008, "Alternative fuels and advanced vehicles data centre", Energy Efficiency and Renewable Energy, US DOE, accessed from www.afdc.energy.gov/afdc, December 2008
4. Anand H, Balasundaram B, Pandit AB, Harrison STL, 2007, "The effect of chemical pretreatment combined with mechanical disruption on the extent of disruption and release of intracellular protein from E. coli", Biochemical Engineering Journal, Vol. 35, pp. 166–173
5. Anthony EJ, 1995, "Fluidised bed combustion of alternative solid fuels: status, successes and problems of the technology", Progress in Energy and Combustion Science, Vol. 21, pp. 239-268.
6. Aracruz Annual and Sustainability report, 2006, accessed from http://www.aracruz.com/minisites/ra2006/en/home/default.htm
7. aracruz.com, 2007, "Generation and consumption of electrical energy and fuels – 2007", accessed from http://www.aracruz.com/minisites/ra2007/, November 2008
8. Baikow VE, 1982, "Sugar series Volume 2: Manufacturing and Refining of Raw Sugar Cane", 2nd completely revised edition, Elsevier, 1982, Amsterdam
9. Bailey JE and Ollis DF, 1986, "Biochemical Engineering Fundamentals", 2nd edition, McGraw-Hill International Editions, Singapore
10. Bain RL, Overend RP, Craig KR, 1998, "Biomass-fired power generation", Fuel Processing Technology, Vol 54, pp. 1-16
11. Baker EG, Mudge LK, Brown MD, 1987, "Steam Gasification of Biomass with Nickel Secondary Catalysts", Industrial and Engineering Chemical Research, Vol. 26, 1335 – 1339.
12. Bane T, 2000, "Sludge disposal strategies", presented at Royal Society of Chemistry sludge treatment seminar, accessed from www.solidstechnology.com, November 2008

13. Barber WP, Stuckey DC, 1999, "The use of the Anaerobic Baffled Reactor (ABR) for wastewater treatment: a review", Water Research, Vol. 33, pp. 1559-1578
14. Barsanti L and Gaultieri P, 2006, "Algae: Anatomy, Biochemistry and Biotechnology", CRC Taylor and Francis Group, USA
15. Beeharry RP, 1996, "Extended sugarcane biomass utilisation for exported electricity production in Mauritius", Biomass and Bioenergy, Vol. 11, pp. 441-449
16. Beeharry RP, 2001, "Strategies for augmenting sugarcane biomass availability for power production in Mauritius", Biomass and Bioenergy, Vol. 20, pp. 421-429
17. Bhatt MS and Rajkumar N, 2001, "Mapping of combined heat and power systems in cane sugar industry, Applied Thermal Engineering, Vol. 21, pp. 1707 - 1719
18. Biljetina R, 1987, "Commercialisation and economics" in Chynoweth DP and Isaacson R (eds) *Anaerobic Digestion of Biomass*, Elsevier Applied Science Publisher, England.
19. Boone D, Mah R, 1987, "Transitional bacteria" in Chynoweth DP and Isaacson R (eds) *Anaerobic Digestion of Biomass*, Elsevier Applied Science Publisher, England
20. Botnia Environmental Balance Sheet, 2006, accessed from www.botnia.com/en, November 2008
21. Boyd RJA, Molema J, Deacon SL, van Niekerk AM, 2008, "Advances in sludge treatment and stabilisation at Johannesburg's wastewater treatment works", presented at the WISA Biennial 2008 Conference, Sun City, South Africa
22. Brage C, Yu Q, Sjostrom K, 1996, "Characteristics of evolution of tar from wood pyrolysis in a fixed-bed reactor", Fuel, Vol. 75, pp. 213-219
23. Bridgewater AV, 2003, "Renewable fuels and chemicals by thermal processing of biomass", Chemical Engineering Journal, Vol. 91, pp. 87-102
24. Bridgewater, AV, 1995, "The technical and economic feasibility of biomass gasification for power generation", Fuel, Vol. 74, pp. 631-653
25. Britz TJ, Meyer LC, Botes PJ, 1983, "Anaerobic digestion of a petrochemical effluent", Biotechnology Letter, Vol 5, pp. 113-118
26. Burton SB, Cohen S, Harrison, STL, Pather-Elias S, Stafford W, van Hille R, von Blottnitz H, 2007, "Energy from wastewater – a feasibility study, Technical report", Water Research Commission Report, Project K5/1732
27. Chamberlain D, Essop H, Hougaard C, Malherbe S, Walker R, 2005a, "Economics of the forestry, timber, pulp and paper industry in South Africa, Part II: Market analysis",

Genesis market report 29 June 2005, accessed from www.forestry.co.za, November, 2008

28. Chamberlain D, Essop H, Hougaard C, Malherbe S, Walker R, 2005b, "Economics of the forestry, timber, pulp and paper industry in South Africa, Part III: Technical notes and appendices", Genesis market report 29 June 2005, accessed from www.forestry.co.za, November 2008
29. Chambers AK, Potter I, 2002, "Gas utilisisation from Sewerage Sludge", Technical Brochure no. 5, published by Carbon and Energy Management, Alberta Research Council, accessed from www.aidis.org.br, January 2009
30. Chirone R, Salatino P, Scala F, Solimene R, Urciuolo M, 2008, "Fluidized bed combustion of pelletized biomass and waste derived fuels", Combustion and Flame, Vol. 155, pp. 21-36
31. Crawford RL, 1981, "Lignin Biodegradation and Transformation", John Wiley and Sons, Canada.
32. Cullis IF, Saddler JN, Mansfield SD, 2004, "Effect of initial moisture content and chip size on the bioconversion efficiency of softwood lignocellulosics", Biotechnology and Bioengineering, Vol. 85, pp. 413 – 421
33. Davidson JF, Harrison D, Darton RC, LaNauze RD, 1977, "The two-phase theory of fluidisation and its application to chemical reactors", in Lapidus L and Amundson NR Chemical Reactor Theory: A Review, Prentice Hall Inc, New Jersey
34. Davidson JF, Harrison D, Darton RC, LaNauze RD, 1977, "The two-phase theory of fluidisation and its application to chemical reactors", in Lapidus L and Amundson NR Chemical Reactor Theory: A Review, Prentice Hall Inc, New Jersey
35. Davidson JF, Harrison D, Darton RC, LaNauze RD, 1977, "Two-phase theory of fluidisation and its application to chemical reactors", in Chemical Reactor Theory: A Review by Lapidus L and Amundson NR, Prentice Hall Inc, New Jersey
36. Deacon S, 2008, Best Practices Manager, Johannesburg Water, Personal communication, October 2008
37. Demirbas A, 2004, "Current technologies for the thermo-conversion of biomass into fuels and chemical", Energy Sources, Vol. 26, pp. 715 – 730
38. Dennis JS, Hayhurst AN, Mackley, IG, 1982, "The ignition and combustion of propane air mixtures in a fluidised bed", International Symposium of Combustion, Combustion Institute, Haifa, Israel

39. Dildar M, 2008, Senior Engineer, eThekwini Water and Sanitation Unit, Personal communication, October 2008
40. DME (Department of Minerals and Energy), 2005, Energy efficiency strategy for the Republic of South Africa, accessed from www.dme.gov.za, October 2008
41. DME, 2003, White paper on the renewable energy policy of the Republic of South Africa, accessed from www.dme.gov.za, October 2008
42. DME, 2008, "South Africa's CDM project portfolio", accessed from www.dme.gov.za, November 2008.
43. dme.gov.za, 2008, "Electricity", accessed from www.dme.gov.za , November 2008
44. dpi.qld.gov.au/environment, 2009, "Feedlot waste management series: Manure production data", Department of Primary Fisheries and Industry, Queensland Government, accessed from www2.dpi.qld.gov.au/environment/5166.html, January 2009
45. Draft White Paper on Water Services, 2002, Department of Water Affairs and Forestry, accessed from www.dwaf.gov.za, October 2008
46. ec.europa.eu, 2001, "Researching the sludge directive", conference report, EC Environment DG and UK Water Industry Research, accessed from www.ec.europa.eu/environment, October 2008
47. ECN 2008, "Phyllis, database for biomass and waste", published by the Energy Research Centre of the Netherlands, accessed from http://www.ecn.nl/phyllis, January 2009
48. Edwards M, 2008, Retired director of Forestry South Africa (FSA), Personal communication, October 2008
49. EEA, 2008, Final energy consumption by sector page, accessed from www.dataservice.eea.europa.eu , November 2008
50. EERE, 2008, "Biomass Energy Data Book", Energy Efficiency and Renewable Energy (EERE), US Department of Energy, accessed from www.cta.ornl.gov/bedb/biofuels, November 2008
51. Eleftheriades C, 2008, Carbon and Environmental Options, Personal communication, November 2008
52. Energy Efficiency and Renewable Energy (EERE), 2008, Biomass feedstock database, accessed from www.eere.energy.gov , November 2008

53. Engler CR, 1985, "Disruption of Microbial Cell" in Moo-Young M (ed) *Comprehensive Biotechnology Vol. 2*, Pergamon Press
54. EPA, 2008, "Sewerage sludge incineration", US Environmental Protection Agency Technology Transfer Network – Clearinghouse for inventories and emissions factors, accessed from http://www.epa.gov , January 2009
55. Eskom Annual report, 2006, accessed from www.eskom.co.za, December 2008
56. EurActiv.com, 2008, "MEPs seek reduced biofuel commitments", press release published on 28 May 2008, accessed from www.euractiv.com, November 2008
57. European Parliament press release, 2008, "More sustainable energy in road transport targets" published on 11 Spetember 2008, accessed from www.europarl.europa.eu, November 2008
58. Fannin KF, Biljetina R, 1987, "Reactor designs" in Chynoweth DP and Isaacson R (eds) *Anaerobic Digestion of Biomass*, Elsevier Applied Science Publisher, England.
59. Fargione J, Hill J, Tilman D, Polasky S, Hawthorne P, 2008, "Land clearing and the biofuel carbon debt", Science, Vol. 319, pp. 1235 – 1238
60. Flickinger MC, Drew SW (eds), 1999, "Encyclopedia of Bioprocess Technology: Fermentation, Biocatalysis and Bioseperation", Vol 1, John Wiley and Sons, New York, pp.137-170
61. Foelkel C, 2007, "Eco-efficient management of the woody forest residues from the eucalyptus plantation forestry", online book accessed from www.eucalyptus.com.br, November 2008
62. Forestry Industry Facts, 2005, published by Forestry South Africa (FSA), accessed from www.forestry.co.za, November 2008
63. Fountain P, 2008, Senior Consultant, Thames Water, Personal communication, February, 2008
64. Fryda LE, Panopoulas KD, Kakaras E, 2007, "Agglomeration in fluidised bed gasification of biomass", Powder Technology, Vol 179, pp. 108-121
65. Fryda LE, Panopoulos KD, Kakaras E, 2007, "Agglomeration in fluidised bed gasification of biomass", Powder Technology, Vol. 179, pp. 108 – 121
66. genesisqa.com, 2008, "Guidelines for the application of sewerage sludge to land", accessed from www.genesisqa.com , November 2008
67. Harrison, STLH, 1991, "Bacterial cell disruption: a key unit operation in the recovery of intracellular products", Biotechnology Advances, Vol 9, pp. 217-245

68. Hayhurst AN, Parmar MS, 1998, "Does solid carbon burn in oxygen to give the gaseous intermediate CO or produce CO2 directly? Some experiments in a hot bed of sand fluidized by air", Chemical Engineering Science. Vol. 53, pp. 427-438
69. Helm R, 2000, "Lignin", Lecture notes on Wood chemistry, products and processes (WOOD 3434), Virginia Polytechnic Institute and State University, accessed from www.woodscience.vt.edu, August 2008
70. Hesketh RP and Davidson JF, 1991, "The effect of volatiles on the combustion of char in a fluidised bed" Chemical Engineering Science, Vol. 46, No. 12, pp.3101-3113
71. Hoffman P, 2008, Utilities Manager Newlands Brewery, SABMiller, Personal communication, October 2008
72. Holl MA, Gush, MB, Hallowes J, Versfeld DB, 2007, "*Jatropha curcas* in South Africa: an assessment of its water use and bio-physical potential", Water Research Commission report no. 1497/1/07
73. Howard M, 2008, Consultant, Fractal Forests, Personal Communication, December 2008
74. IEA, 2004, "Biofuels for transport – An International Perspective", International Energy Agency (IEA), accessed from www.iea.org, November 2008
75. IFPRI, 2007, Food policy report entitled "The world food situation: new driving forces and required action" by Joachim von Braun, accessed from www.ifpri.org, November 2008
76. IFPRI, 2008b, testimony to US Senate, "Biofuels, international food prices and the poor", by Joachim von Braun, accessed from www.ifpri.org, November 2008
77. International Food Price Research Institute (IFPRI), 2008a, presentation entitled "Food prices, biofuels and climate change", accessed from www.ifpri.org, November 2008
78. Jeffries T, 1997, "Kraft pulping – Energy Consumption and Production", lectures notes accessed from http://calvin.biotech.wisc.edu, January 2009
79. Jenkins BM, Baxter LL, Miles Jr. TR, Miles TR, 1998, "Combustion properties of biomass", Fuel Processing Technology, Vol. 54, pp.17–46
80. Kelly H G, 2006, "Emerging processes in biosolids treatment, 2005", Journal of Environmental Engineering and Science, Vol. 5, pp. 175 – 186
81. Kerr I, 2008, Senior Research Fellow, Paper Manufacturers Association of South Africa (PAMSA), Personal communication, October 2008

82. Kiel JHA (ed), 2004, "Primary measures to reduce tar formation in fluidised bed biomass gasifiers", Final report SDE project P1999-012, ECN
83. Kim J, Park C, Kim T-H, Lee M, 2003, Effects of Various Pretreatments for Enhanced Anaerobic Digestion with Waste Activated Sludge", Journal of Bioscience and Bioengineering, Vol. 95, pp. 271 – 275
84. King P, 2008, Cape Town Water and Sanitation Services, Personal Communication, September 2008
85. Klimantos P, Koukouzas N, Katsiadakis A, Kakaras E, 2008, "Air-blown biomass gasification combined cycles (BGCC): System analysis and economic assessment", Energy, in press, doi:10.1016/j.energy.2008.04.009
86. Koornneef J, Junginger M and Faaij A, 2007, "Development of fluidized bed combustion - An overview of trends, performance and cost", Progress in Energy and Combustion Science, Vol 33, pp. 19 – 55.
87. La Nauze RD, Duffy GJ, 1985, "Coal rejects – a wasted resource?", Environmental Geochemistry and Health, Vol 7, pp. 69 – 79.
88. Liebenberg AJ, 2008, Senior Researcher, Agricultural Research Council, Personal Communication, September 2008
89. Lynd LR, von Blottnitz H, de Boer J, Tate B, Pretorius IS, Rumbold K, van Zyl WH, 2003, "Plant Biomass Conversion to Fuels and Commodity Chemicals in South Africa: A Third Chapter?", South African Journal of Science, Vol. 99, pp. 499-507
90. Magasiner N, 1987, "The effect of fuel moisture content on the performance of a typical bagasse fired watertube boiler", Proceedings of the South African Sugar Technologists' Association, Vol. 60, pp. 86 – 89
91. Magasiner N, Van Alphen C, Inkson MB, Misplon BJ, 2001, Characterising fuels for biomass-coal co-fired generation, Proceedings of the South African Sugar Association, Vol. 75, pp. 282 – 291
92. Marrison CI, Larson ED, 1996, "A preliminary analysis of the biomass energy production potential in Africa in 2025 considering projected land needs for food production", Biomass and Bioenergy, Vol. 10, pp. 337-351
93. McCabe W L, Smith J C and Harriott P, 2005, "Unit operations of Chemial Engineering", 7th edition, International edition, McGraw-Hill, Singapore
94. McKendry P, 2000a, "Energy from biomass (part 1): overview of biomass", Bioresource Technology Vol. 83, pp. 37–46

95. McKendry P, 2000b, "Energy production from biomass (part 2): conversion technologies", Bioresource Technology, Vol. 83, pp.47–54
96. McKendry P, 2002c, "Energy production from biomass (part 3): gasification technologies", Bioresource Technology, Vol. 83 pp.55–63
97. Middelberg APJ, 1995, "Process scale disruption of microorganisms", Biotechnology Advances, Vol. 13, pp. 491 – 551
98. Milne TA, Evans RJ, 1998, "Biomass Gasifier "Tars": Their Nature, Formation, and Conversion", prepared for National Renewable Energy Laboratory, report reference: NREL/TP-570-25357
99. Mohee R and Beeharry RP, 1999, "Life cycle analysis of compost incorporated sugarcane bioenergy systems in Mauritius", Biomass and Bioenergy, Vol. 17, pp. 73-83
100. Mosier N, Wyman C, Dale B, Elander R, Lee YY, Holtzapple M, Ladisch M, 2005, "Features of promising technologies for pretreatment of lignocellulosic biomass", Bioresource Technology, Vol. 96, pp. 673–686
101. Nah, IW, Kang YW, Hwang K-Y, Woong KS, 2000, "Mechanical Pretreatment of Waste Activated Sludge for Anaerobic Digestion Process", Water Research, Vol. 34, pp. 2362-2368
102. Narvaez I, Orio A, Aznar MP, Corella J, 1996, "Biomass Gasification with Air in an Atmospheric Bubbling Fluidized Bed. Effect of Six Operational Variables on the Quality of the Produced Raw Gas", Industry, Engineering and Chemistry Research, Vol. 25. pp. 2110 – 2120.
103. National Business Initiative (NBI), 2008, accessed from www.nbi.org.za, November, 2008
104. Natsurv 1, 1987, Water and Wastewater Management in the Malt Brewing Industry, Water Research Commision report, report no. TT29/87
105. Natsurv 11, 1990, Water and Wastewater Management in the Sugar Industry, Water Research Commision report, report no. TT47/90
106. Natsurv 12, 1990, Water and Wastewater Management in the Pulp and Paper Industry, Water Research Commision report, report no. TT49/90
107. Natsurv 14, 1990, Water and Wastewater Management in the Wine Industry, Water Research Commision report, report no. TT51/90
108. Natsurv 15, 2005, Water and Wastewater Management in the Oil Refining and Re-refining Industry, Water Research Commision report, report no. TT80/05

109. Natsurv 3, 1987, Water and Wastewater Management in the Soft Drink Industry, Water Research Commision report, report no. TT35/87
110. Natsurv 4, 1989, Water and Wastewater Management in the Dairy Industry, Water Research Commision report, report no. TT38/89
111. Natsurv 5, 1989, Water and Wastewater Management in the Sorghum Malt Beer Industry, Water Research Commision report, report no. TT39/89
112. Natsurv 6, 1989, Water and Wastewater Management in the Edible Oil Industry, Water Research Commision report, report no. TT40/89
113. Natsurv 7, 1989, Water and Wastewater Management in the Red Meat Industry, Water Research Commision report, report no. TT41/89
114. Natsurv 9, 1989, Water and Wastewater Management in the Poultry Industry, Water Research Commision report, report no. TT43/89
115. Nel A, 2008, Senior Researcher, Agricultural Research Council, Personal Communication, September 2008
116. Neyens E, Baeyens J, 2003, A review of thermal sludge pre-treatment processes to improve dewaterability", Journal of Hazardous Materials Vol. B98, pp. 51–67
117. Nijenhuis J, Korbee R, Lensselink J, Kiel JHA, van Ommen JR, 2007, "A method for agglomeration detection and control in full-scale biomass fired fluidized beds", Chemical Engineering Science, Vol 62, pp. 644 - 654
118. Noike T, Endo G, Chang J, Yaguchi J, Matsumoto J, 1985, "Characteristics of carbohydrate degradation and the rate-limiting step in anaerobic digestion", Biotechnology and Bioengineering, Vol. 28, pp. 1482 – 1489
119. Nordgreen T, Liliedahl T, Sjostrom K, 2006, "Metallic iron as a tar breakdown catalyst related to atmospheric, fluidised bed gasification of biomass", Fuel, Vol. 85, pp. 689–694
120. North BC, Engelbrecht AD, Hadley TD, 2005, "Co-firing coal and biomass in a fluidised-bed boiler", Proceedings of the International Fluidisation South Africa conference, Johannesburg, South Africa
121. Ntsowe D, 2008, Wastewater Treatment, City of Tshwane, Personal communication, October 2008
122. Nussbaumer T, 2003, "Combustion and Co-combustion of Biomass: Fundamentals, Technologies, and Primary Measures for Emission Reduction", Energy and Fuels, Vol. 17, pp. 1510-1521

123. Ogada T and Werther J, 1996 "Combustion characteristics of wet sludge in a fluidized bed-Release and combustion of the volatiles", Fuel, Vol 75, No. 5, pp. 617-626
124. Ohman M, Nordin A, 2000, "Bed Agglomeration Characteristics during Fluidized Bed Combustion of Biomass Fuels", Energy and Fuels, Vol. 14, pp.169-178
125. Onyeche TI, Schäfer S, 2003, "Sludge homogenisation as a means to reduce sludge volume and increase energy production", Electronic Journal of Environmental, Agriculture and Food chemistry", Vol. 2, pp. 291-296
126. OPSI, 1994, "The Urban Waste Water Treatment (England and Wales) Regulations 1994", accessed from Office of Public Sector Information website, accessed from www.opsi.gov.uk, November 2008
127. Owen WF, Stuckey DC, Healy Jnr JB, Young LY, McCarty PL, 1979, "Bioassay for monitoring biochemical methane potential and anaerobic toxicity", Water Research, Vol. 13, pp. 485 – 492
128. Paterson, N, 1997, "Fuel behaviour studies in the air-blown gasification cycle", Fuel, Vol. 76, pp. 1319 – 1325
129. Perry RH and Green DW, Maloney JO (eds), 1999, "Perry's Chemical Engineers Handbook", 7th Edition, McGraw-Hill, USA
130. Piterou A, Shackley S, Upham P, 2008, "Project ARBRE: Lessons for bio-energy developers and policy-makers", Energy Policy, Vol. 36, pp. 2044– 2050
131. Purchase BS, Wynne AT, Meyer E, Van Antwerpen R, 2008, "Is there profit in cane trash ? - Another dimension to the assessment of trashing versus burning". Proceedings of the South African Sugar Technologists' Association, Vol. 81, pp. 86-99
132. Quaak P, Knoef H, Stassen H, 1999, "Energy from Biomass – A review of Combustion and Gasification Technologies", World Bank Technical Paper No. 422, Energy Series, Washington DC
133. Rapagna S, Jand N, Kiennemann A, Foscolo PU, 2000, "Steam gasifcation of biomass in a fuidised-bed of olivine particles", Biomass and Bioenergy, Vol. 19, pp. 187 -197
134. Rawlinson J, 2008, Engineering manager at Maidstone Sugar mill, Tongaat-Hulett, Personal communication, October 2008
135. REN21, 2007, "Renewables 2007: global status report", accessed from www.ren21.net, November 2008

136. Rodrigues M, Walter A, Faaij A, 2003, "Co-firing of natural gas and Biomass gas in biomass integrated gasification/combined cycle systems", Energy, Vol. 28 pp.1115–1131
137. Rosillo-Calle F, Bajay SV, Rothman H (eds.), 2000, "Industrial uses of biomass energy: The example of Brazil", Taylor and Francis, Great Britain
138. Saha BC, 2001, "Hemicellulose bioconversion", Industrial Microbiology and Biotechnology, 2003, Vol. 30, pp. 279–291
139. SASA, 2008, South African Sugar Association crop data, obtained from SASA communications officer (Jennene Singh)
140. Schafer KA, Wallis FR, 2005, "Livestock manure production rates and nutrient content", 2005 North Carolina Agricultural Chemicals Manual, College of Agricultural and Life Science, NC State University, accessed from www.lib.clemson.edu/agnic, October 2008
141. Schink B, 2008, "Energetic aspects of methanogenic feeding webs", in Wall JD, Harwood CS, Demain A (eds) *Bioenergy*, ASM Press, Washington USA
142. Schuster G, Löffer G, Weigl K, Hoffbauer H, 2001, "Biomass steam gasification – an extensive parametric modeling study", Bioresource Technology, Vol 77, pp. 71-79.
143. Schwarz WH, 2001, "The cellulosome and cellulose degradation by anaerobic bacteria", Applied Microbiology Biotechnology, Vol 56, pp. 634–649
144. Scott SA, 2004, "The Gasification and Combustion of Sewage Sludge in a Fluidised Bed", PhD dissertation, Chemical Engineering, University of Cambridge
145. Shana A, 2005, "Reading Sewerage Treatment Works Performance – monitorinig report", Thames Water internal report
146. Shana A, 2006, "Performance of the Chertsey Digestion Plant with CAMBI™ thermal hydrolysis pretreatment", Thames Water internal report
147. Skrifvars B-J, Ohman M, Nordin M, Hupa M, 1999, "Predicting Bed Agglomeration Tendencies for Biomass Fuels Fired in FBC Boilers: A Comparison of Three Different Prediction Methods", Energy and Fuels, Vol. 13, pp.359-363
148. Skrifvars B-J, Ohman M, Nordin M, Hupa M, 1999, "Predicting Bed Agglomeration Tendencies for Biomass Fuels Fired in FBC Boilers: A Comparison of Three Different Prediction Methods", Energy and Fuels, Vol. 13, pp.359-363
149. Sleat R and Mah R, 1987, "Hydrolytic bacteria" in Chynoweth DP and Isaacson R (eds) *Anaerobic Digestion of Biomass*, Elsevier Applied Science Publisher, England

150. Snyman HG and Herselman JE, 2006a, “Guidelines for the Utilisation and Disposal of Wastewater Sludge Volume 2: Requirements for the agricultural use of wastewater sludge”, WRC report no. TT 262/06
151. Snyman HG and Herselman JE, 2006b, “Guidelines for the Utilisation and Disposal of Wastewater Sludge Volume 1: Selection of Management options”, WRC report no. TT 261/06
152. Snyman HG, Herselman JE, Kasselman G, 2004, A metal content survey of South African sludges and an evaluation of analytical methods for their determination in sludge”, Water Research Commission report, report no. 1283/1/04
153. Speece RE, 1996, “Anaerobic Biotechnology for Industrial Wastewaters”, Vanderbilt University, Archae Press, USA
154. Stephenson AL, Dennis JS, Scott SA, 2008, “Improving the sustainability of the production of biodiesel from oilseed rape in the UK”, Process Safety and Environment Protection, in press, doi:10.1016/j.psep.2008.06.005
155. Stephenson R, 2006, “Demonstration of the MicroSludge Process to Pre-treat WAS at Los Angeles County Joint Water Pollution Control Plant”, presented at BC Water and Waste Association 2006 Conference and Trade Show April 29 – May 3, 2006 Whistler, BC, accessed from www.bcwwa.org, November 2008
156. Stephenson RJ, Laliberte S, Hoy PM, Britch D, 2007, Full scale and laboratory scale results from the trial of MicroSludge at the Joint Water Pollution Control Plant at Los Angeles County”, Water Science, Vol. 1, pp. 1 - 13
157. Strachan LJ, Wright M, Pass J, Couth RJ, Pearson G, Cooke S, 2008, “Great Expectations: Landfill gas production theory vs. reality”, Proceedings of the 19th Waste Management Conference of the IWMSA (WasteCon 2008), 6-10th October, Durban, South Africa
158. Sun Y, Cheng J, 2002, “Hydrolysis of lignocellulosic materials for ethanol production: a review”, Bioresource Technology, Vol 83 pp. 1–11
159. Tait B, 2006, “Sasol’s activities on synthetic fuels”, Proceedings of the 2nd International Biomass-to-Gas Congress, 13 October 2006, Berlin, Germany
160. Tauschitz J, Hochfellner M, 2004, “CO_2- reduction targets call for applying Best Available Technology; a new 800MW combined cycle power plant south fo Graz”, Proceedings of IPCC conference on Energy Efficiency in IPCC installations, Vienna, 21 to 22 October, 2004

161. Tchobanoglous G, Burton FL, Stensel HD (eds), 2003, "Wastewater engineering, Treatment and Reuse", 4th Edition, Metcalf and Eddy Inc, McGraw-Hill, USA
162. Teymouri F, Laureano-Perez L, Alizadeh H, Dale, BE, 2005, "Optimization of the ammonia fiber explosion (AFEX) treatment parameters for enzymatic hydrolysis of corn stover", Bioresource Technology Vol. 96, pp. 2014–2018
163. thameswater.co.uk, 2008, "Our business: Then and now", accessed from www.thameswater.co.uk, November 2008
164. Tongaat-Hulett 2007 Annual report, 21 February 2008, accessed from www.tongaat.co.za , November 2008
165. tonto.eia.doe.gov, 2008, "Energy Information Administration – official energy statistics from the US government", accessed from http://tonto.eia.doe.gov/oog/info/twip/twip_gasoline.html , December 2008
166. tps.se, 2007, access from www.tps.se, January 2009
167. Tsao, GT, 1987, "Pre-/posttreatments", in Chynoweth DP and Isaacson R (eds) *Anaerobic Digestion of Biomass*, Elsevier Applied Science Publisher, England
168. Turnbull E and Davidson JF, 1984, "Fluidized bed combustion of char and volatiles from coal", AIChE Journal, Vol. 30, pp. 881 – 889
169. Turnbull E, 1983, "The fluidized bed combustion of char and volatiles from coal", PhD dissertation, Cambridge University.
170. Ullmans' Encyclopedia of Industrial Chemistry, 2003a, s.v. "Wastewater", Vol 38, 6th edition, Wiley-VCH, Weinheim
171. Ullmans' Encyclopedia of Industrial Chemistry, 2003b, s.v. "Paper and pulp", Vol 25, 6th edition, Wiley-VCH, Weinheim
172. Ullmans' Encyclopedia of Industrial Chemistry, 2003c, s.v. "Sugar", Vol 34, 6th edition, Wiley-VCH, Weinheim
173. UNEP, 2008, "Global trends in sustainable energy investment 2008", accessed from www.unep.fr, November 2008
174. USAID, 2005, "Case studies of bankable water and sewerage utilities Volume 1: Overview report", August 2005, accessed from waterwiki.net, November 2008
175. Valo A, Carrere H, Delgenes JP, 2004, "Thermal, chemical and thermo-chemical pre-treatment of waste activated sludge for anaerobic digestion", Journal of Chemical Technology and Biotechnology, Vol. 79, pp. 1197 - 1203

176. Van Antwerpen R, Meyer JH, Thompson GD, 2006, "The impact of trashing on yield response in the South African Sugar Industry: A summary of results from several BT trials", Proceedings of the South African Sugar Technologists' Association, Vol. 80
177. Van Antwerpen R, Meyer, JH, Turner PET, 2001, "The effects of cane trash on yield and nutrition from the long-term field trial at Mount Edgecombe", Proceedings of the South African Sugar Technologists' Association, Vol. 75, pp. 235 – 241
178. Van der Vaart, DR, 1988 "Freeboard ignition of premixed hydrocarbon gas in a fluidized bed", Fuel, Vol 67, pp. 1003
179. Van Dyk JC, Waanders FB, 2007, "Manipulation of gasification coal feed in order to increase the ash fusion temperature of the coal enabling the gasifiers to operate at higher temperatures", Fuel, Vol 86, pp. 2728-2735
180. Vesilind, PA, 1994, "Sludge dewatering: why the water will win every time", Proceedings of the 49th Industrial Waste Conference, Purdue University
181. von Blottnitz H, Austin G, Nissing C, Schmalbein N, Liphoto L, Ncwadi N, Gets A, Fedorsky C, 2006, "Burn, Gasify, Pyrolyse or Ferment? Making sense of the many possibilities for energy from waste in South Africa", Proceedings of the 18th Waste Management Conference of the IWMSA (WasteCon 2006), 4th – 8th September, Cape Town, South Africa
182. Vymazal J, 1995, "Algae and Element Cycling in Wetlands", Lewis Publishers, USA
183. Waites MJ, Morgon NL, Rockey JS, Higton G, 2001, "Industrial biotechnology: An Introduction", Blackwell Science, Oxford
184. Waldheim L and Carpentieri, 2001, Update on the Progress of the Brazilian Wood BIG-GT Demonstration Project", Journal of Engineering for Gas Turbines and Power, ASME, Vol 123, pp. 525-536
185. Wang L, Weller C L, Jones D D, Hanna M A, 2008, "Contemporary issues in thermal gasification of biomass and its application to electricity and fuel production", Biomass and Bioenergy, Vol. 32 pp. 573 – 581
186. Wang LK, Shammas NK, Hung Y, 2007, "Handbook of Environmental Engineering Vol 6: Biosolids Treatment Processes", Humana Press, New Jersey
187. Weesmaes MPJ and Verstaete WH, 1998, "Review: Evaluation of Current Wet Sludge Disintegration Techniques", Journal of Chemical Technology and Biotechnology, 1998, Vol. 73, pp. 83-92

188. Werther J and Ogada T, 1999, “Sewage sludge combustion”, Progress in Energy and Combustion Science Vol. 25 pp.55–116
189. Werther J, Ogada T, 1999, “Sewage sludge combustion”, Progress in Energy and Combustion Science, Vol. 25 pp.55–116
190. Wilkie AC, 2008, “Biomethane from biomass, biowaste and biofuels”, in Wall JD, Harwood CS, Demain A (eds) *Bioenergy*, ASM Press, Washington USA
191. Winter J (ed), 1999, “Biotechnology, A Multi-Volume Comprehensive Treatise, Volume 11a, Environmental Processes I”, 2nd completely revised edition, Wiley VCH, Germany
192. Wyman CE, Dale BE, Elander RT, Holtzapple M, Ladisch MR, Lee YY, 2005a, “Coordinated development of leading biomass pretreatment technologies”, Bioresource Technology Vol. 96, pp. 1959–1966
193. Wyman CE, Dale BE, Elander RT, Holtzapple M, Ladisch MR, Lee YY, 2005b, “Comparative sugar recovery data from laboratory scale application of leading pretreatment technologies to corn stover”, Bioresource Technology Vol. 96, pp. 2026–2032
194. Zevenhoven-Onderwater M, Backman R, Skrifvars B-J, Hupa M, 2001, “The ash chemistry in fluidised bed combustors of biomass fuels. Part: 1 – Predicting the chemistry of melting ash and ash-bed material interaction”, Fuel, Vol. 80, pp. 1489 – 1502

Appendix A Experimental work appendix

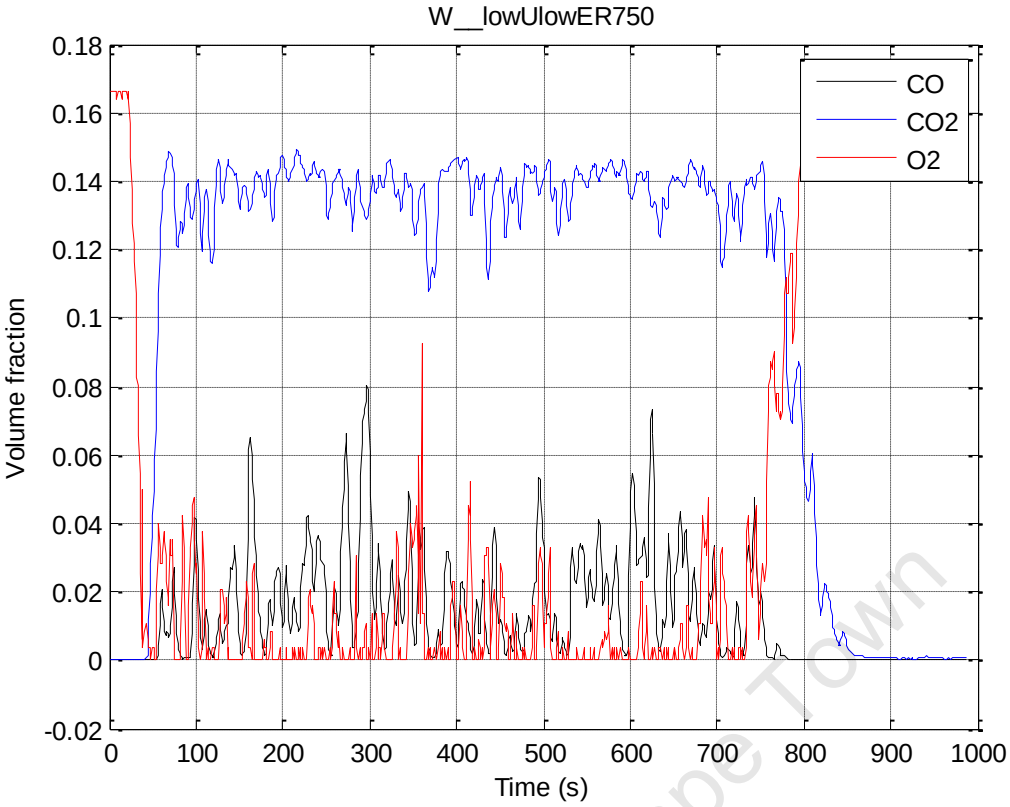
A.1. Run information table

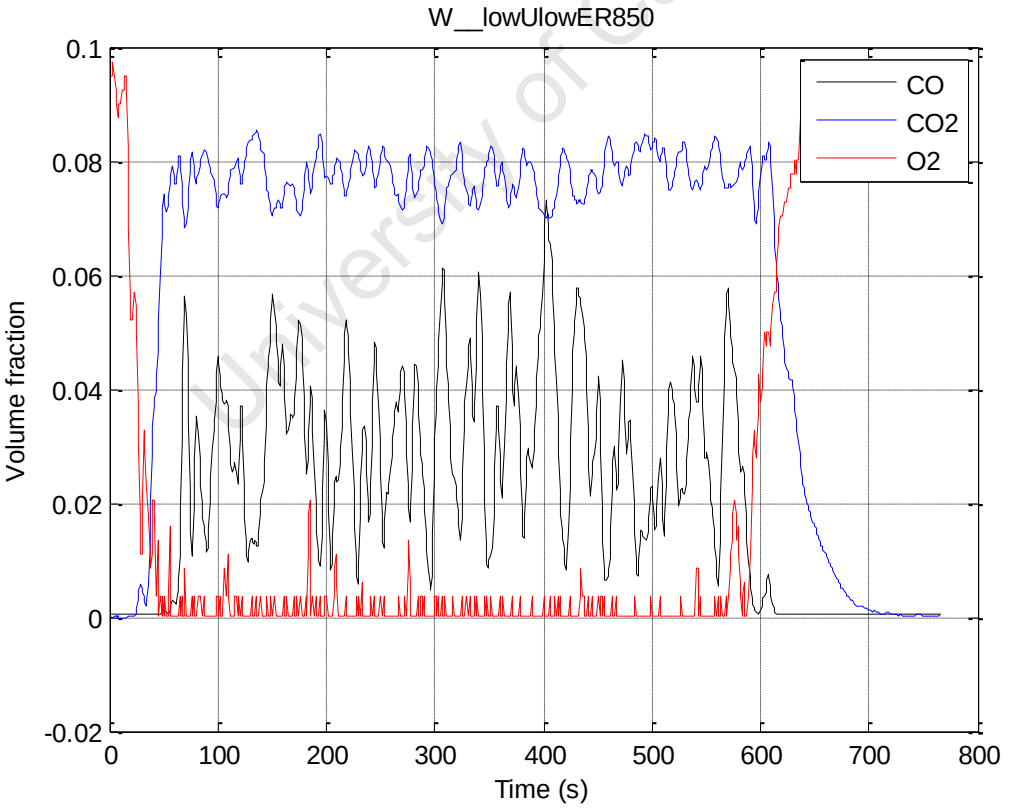
Table A-0-1: Run information

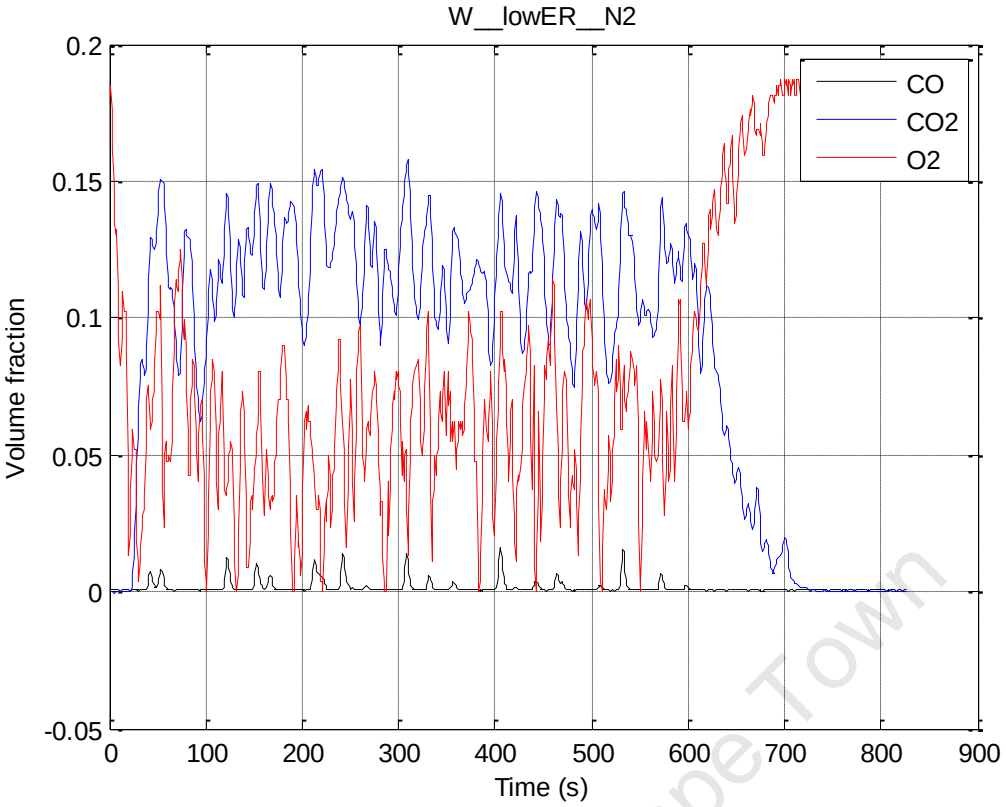
Run number	Run ID	Fuel	Initial bed temperature (°C)	Fuel flowrate (g/min)	U/Umf	Equivalence ratio (ER)
1	'W__BC700'	Wood	714	2.2	6.5	2.0
2	'W__BC750'	Wood	757	2.4	7.1	1.9
3	'W__BC800'	Wood	805	2.4	7.7	1.9
4	'W__BC850'	Wood	866	2.5	8.4	1.8
5	'W__BC2__800'	Wood	805	2.0	7.1	2.0
6	'W__BC2__850'	Wood	860	2.1	7.4	1.9
7	'W__BC3__700'	Wood	709	2.1	6.4	2.2
8	'W__BC3__750'	Wood	750	2.2	6.9	2.0
9	'W__BC3__850'	Wood	798	2.3	7.5	2.0
10	'W__lowUlowER750'	Wood	770	2.3	4.3	1.1
11	'W__lowUlowER850'	Wood	838	2.9	4.8	0.9
12	'W__lowER__N2'	Wood	741	3.5	7.3	0.7
13	'W__lowER'	Wood	746	3.1	6.7	1.4
14	'W__OB750'	Wood	763	2.1	7.1	2.1
15	'W__OB850'	Wood	849	2.2	8.2	2.0
16	'W__OB__lowER'	Wood	754	4.2	7.0	1.1
17	'S__750__O2'	Sludge	760	2.3	6.2	1.5
18	'S__BC700'	Sludge	715	3.8	6.7	1.1
19	'S__BC750'	Sludge	759	3.3	6.9	1.2
20	'S__CB800'	Sludge	800	3.8	7.0	1.0
21	'S__BC850'	Sludge	848	3.2	7.2	1.1
22	'S__lowER700'	Sludge	710	6.3	6.6	0.7

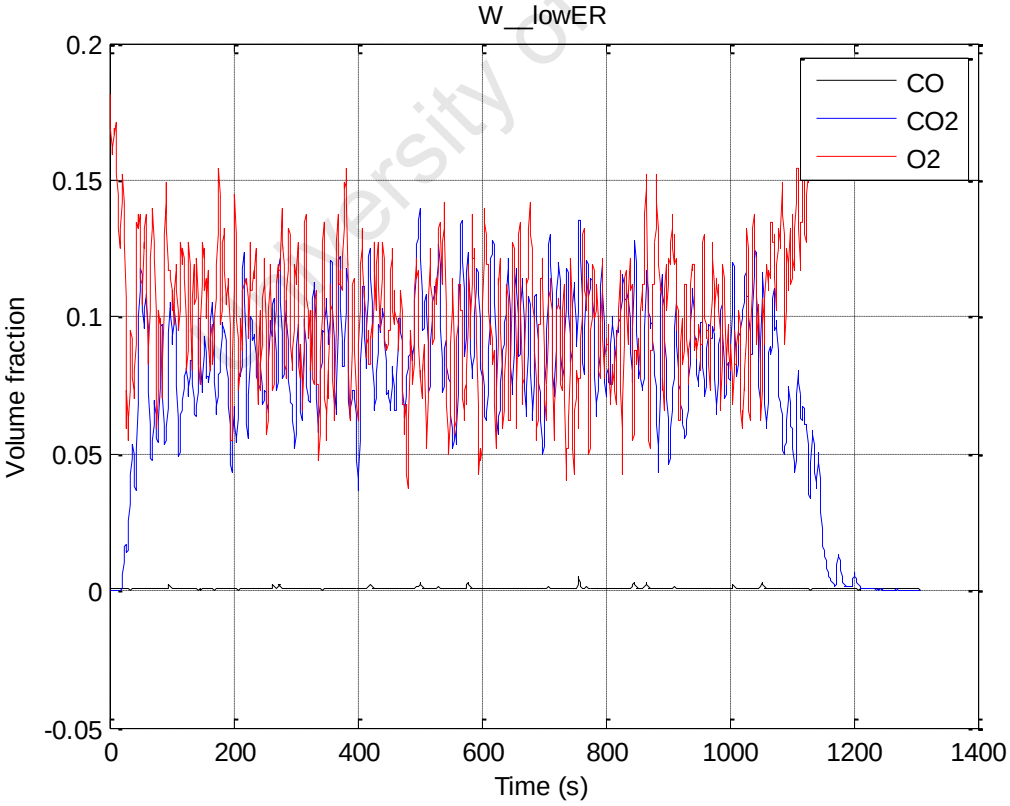
A.2. Mass balance data

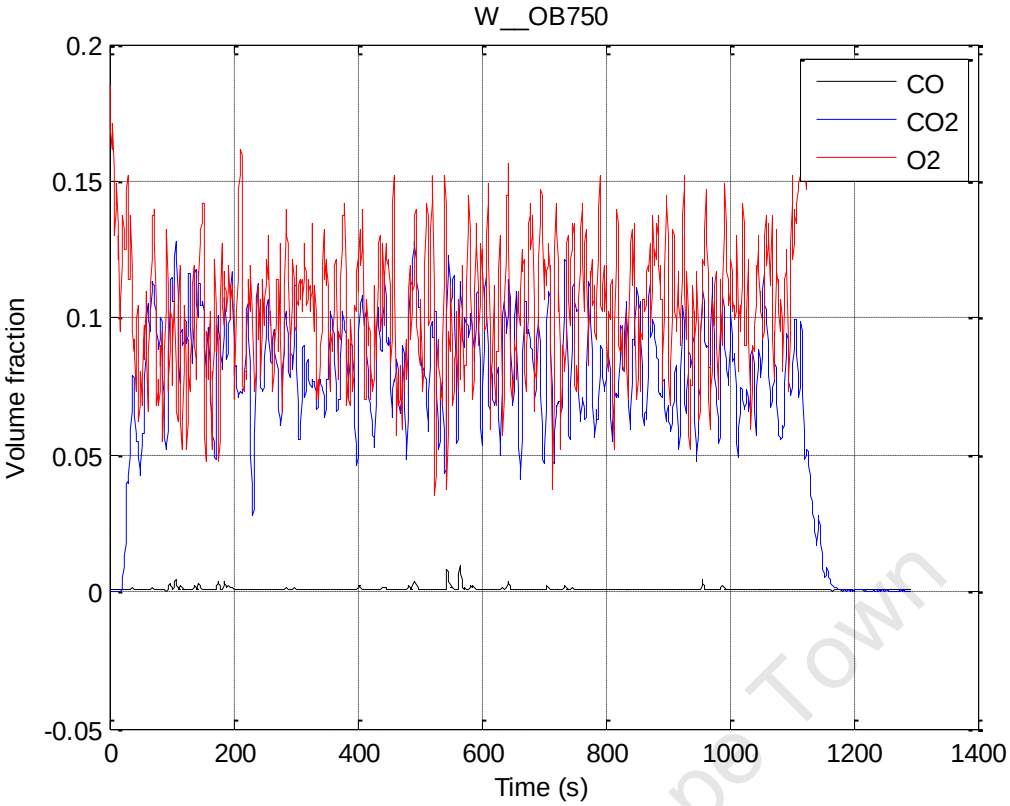
The following plots show the CO_2, CO and O_2 traces for selected runs where the measurement of all three took place. The run ID (as given in Table A-0-1) is given at the top of each plot.

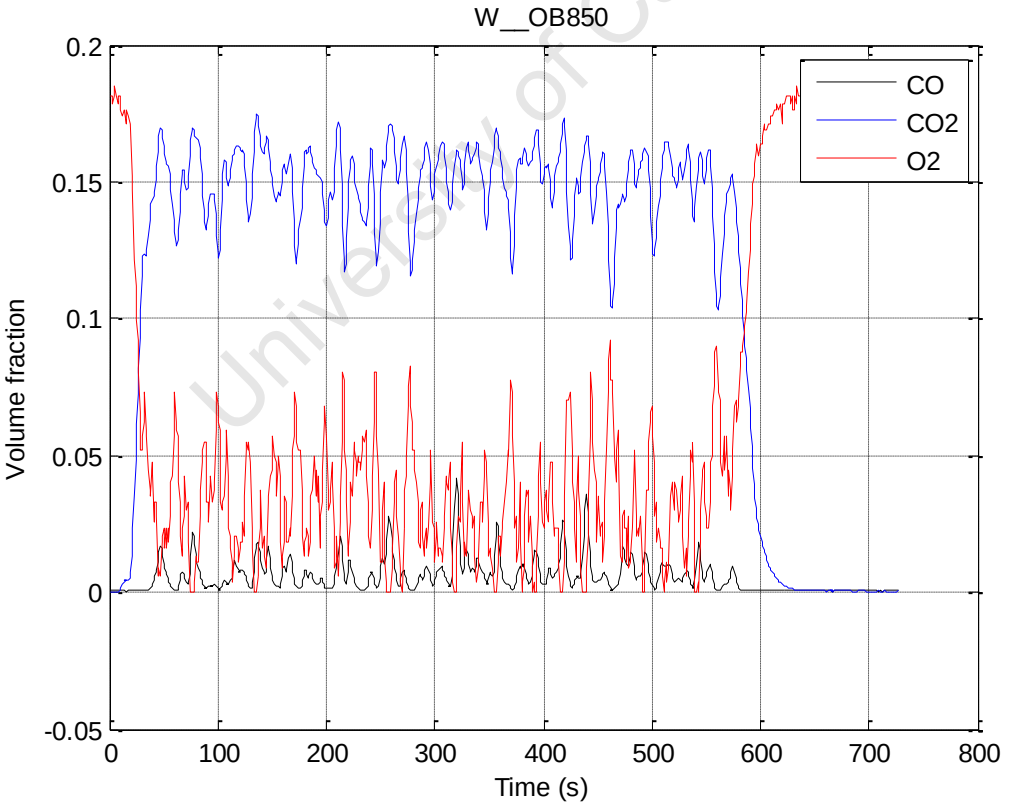


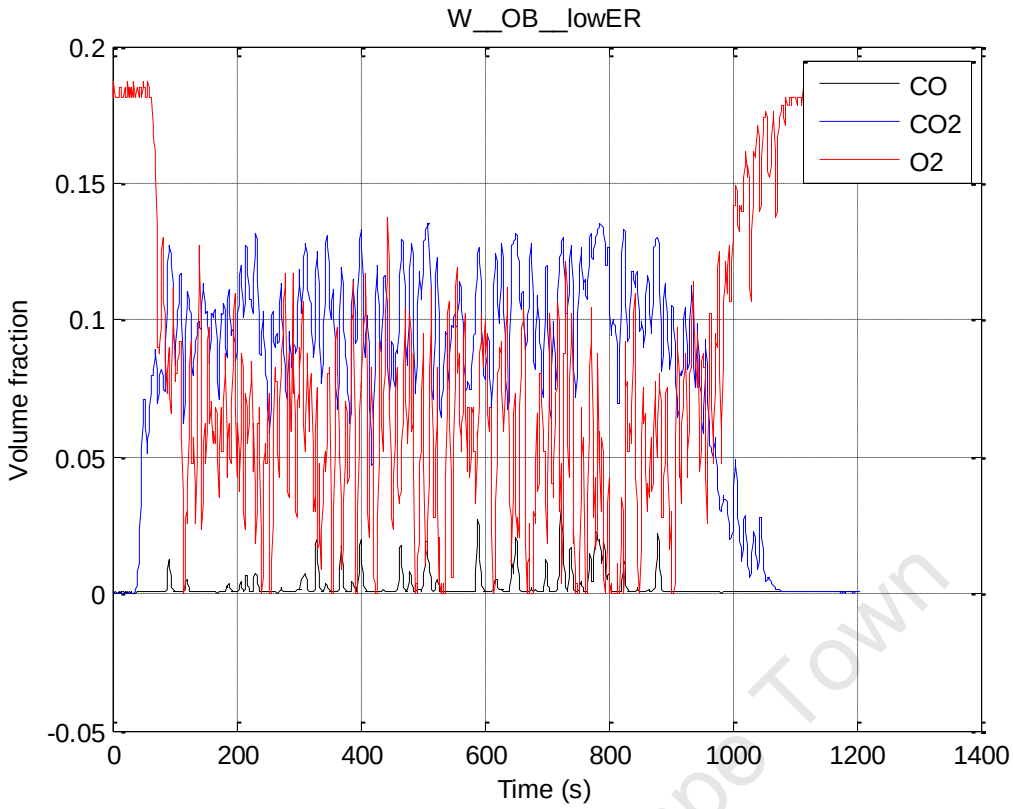












A.3. Energy balance data

A.3.1. Thermodynamic data inputs

The thermodynamic inputs to the the equations of Section 4.2 are shown in Table A-0-2.

Table A-0-2: Thermodynamic inputs

Parameter	Units	Value/Equation	Reference
Cp_{air}	J/kg.K	$Cp = 7E\text{-}05 \cdot T^2 + 0.084 \cdot T + 927.53$	Fitted from Perry's data
Cp_{steel}	J/kg.K	$Cp = 3E\text{-}07 \cdot T^3 - 0.0008 \cdot T^2 + 0.7952 \cdot T + 294.48$	Fitted from Perry's data
$Cp_{alumina}$	J/kg.K	870	Manufacturer's data
ρ_{steel}	kg/m^3	8000	Manufacturer's data
LHV_{wood}	MJ/kg	14	Phyllis database (Phyllis.nl/en)
LHV_{sludge}	MJ/kg	19	Pers. comm.. Thames Water

A.3.2. Heat input calculations

The calculations of the electrical heat input from the furnace, the approximated element temperature and the heat loss through the firebricks are shown in

Table A-0-3: Calculation of electrical heat input from the furnace

Calculation of Q_{in}		
Average %power	%	55%
I	A	22
V	V	240
Q_{in}	W	3000

Table A-0-4: Calculation of furnace element temperature

Calculation of $T_{element}$		
σ	$W/m^2/K^4$	5.67E-08
ε		0.7
A_h per element	m^2	0.013
$Q_{radiative}$ (per heating rod)	W	880
$T_{element}$	K	1139

Table A-0-5: Calculation of heat loss through firebricks

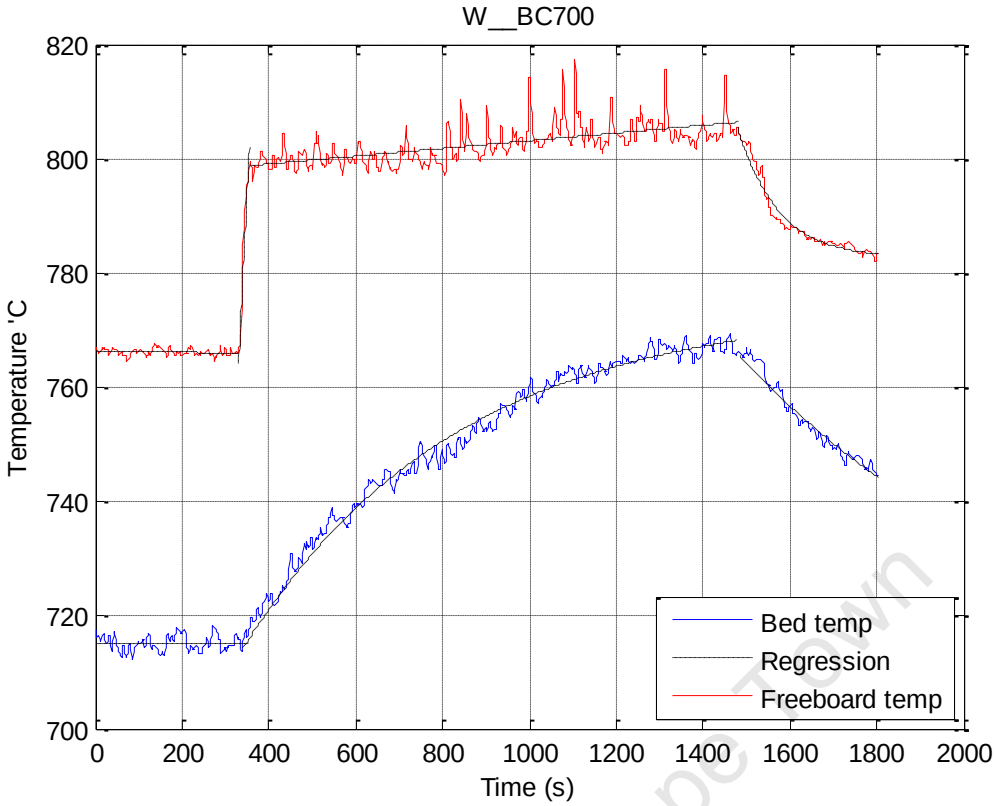
Calculation of Q_{loss}		
$T_{wall,outer}$	K	330
$T_{wall,inner}$	K	1100
k	W/m.K	0.3
l	m	0.15
A_h	m^2	1.6
Q_{loss}	W	2400

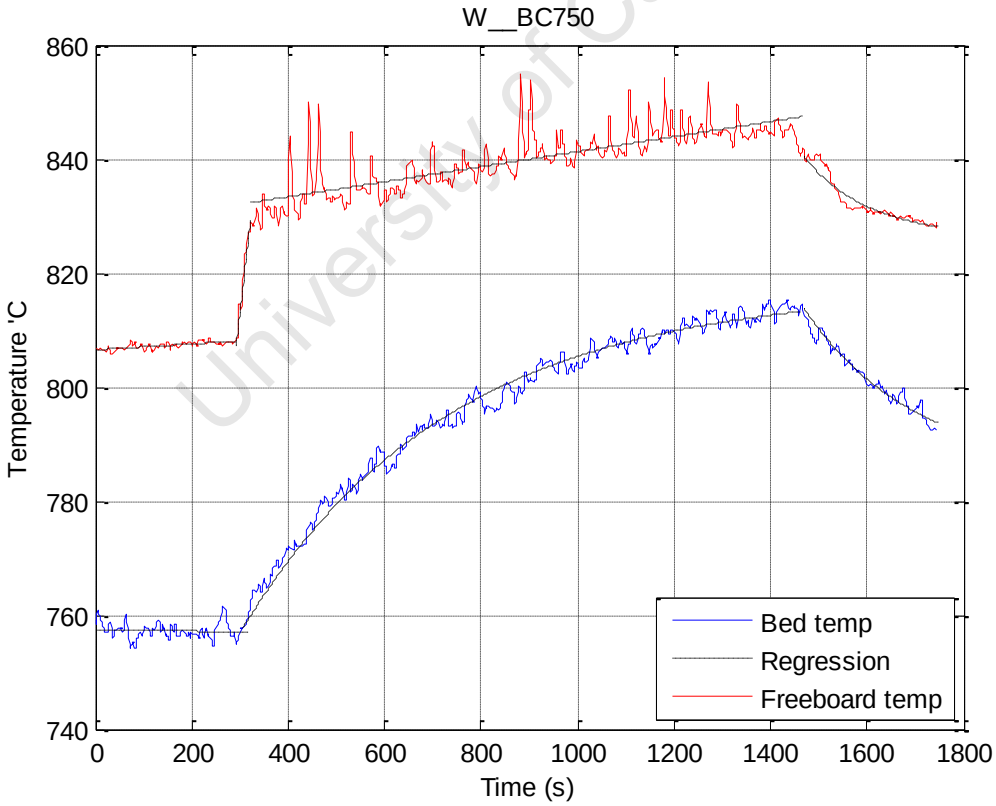
Table A-0-6: Calculation of Biot number

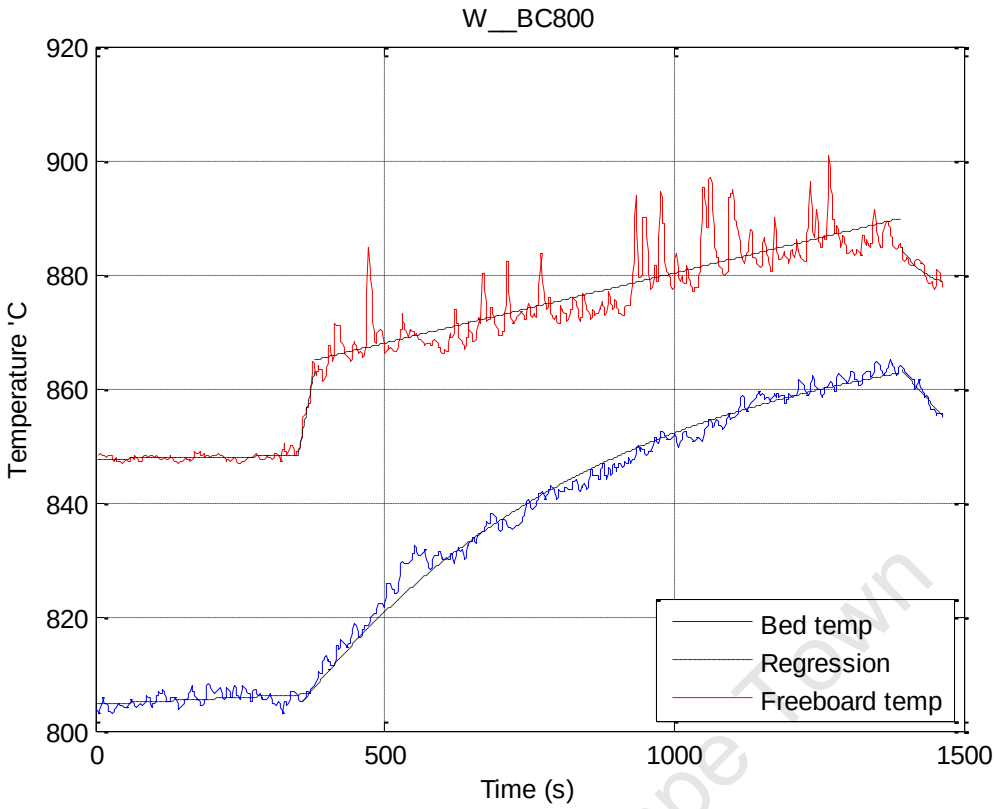
Convective heat transfer coefficient ($h_{bed-wall}$)	$W/m^2.K$	500
Wall conductivity (k)	W/m.K	21
Wall thickness (l)	m	0.006
Biot number	-	0.1

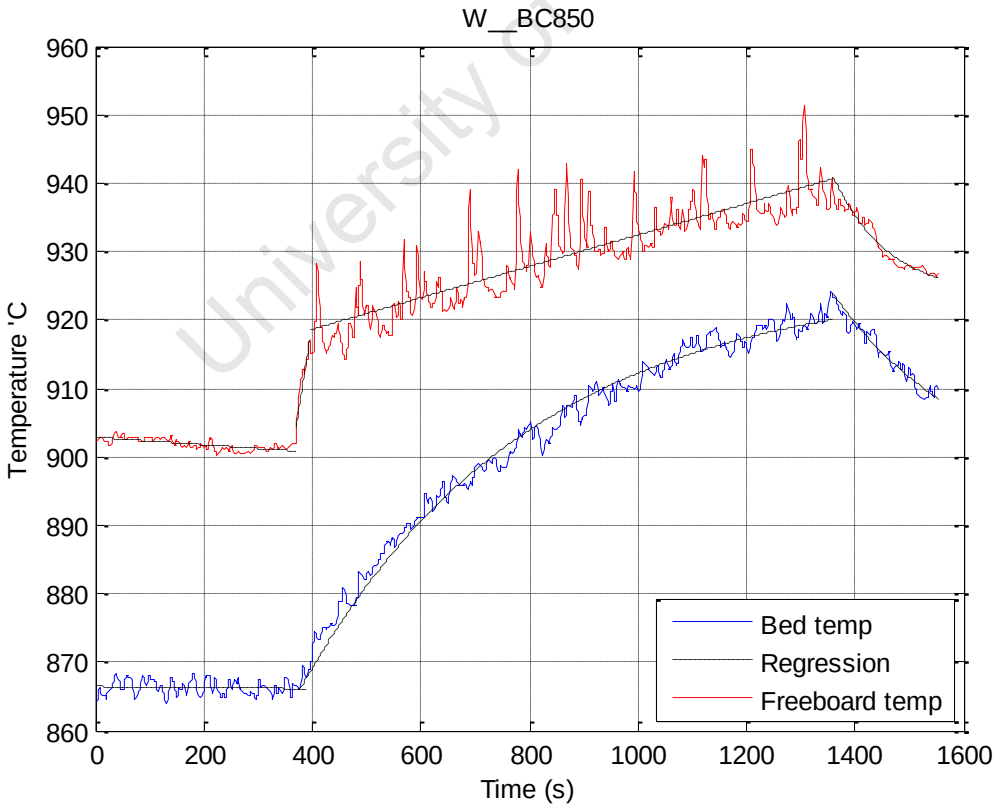
A.3.3. Bed and freeboard temperature raw data

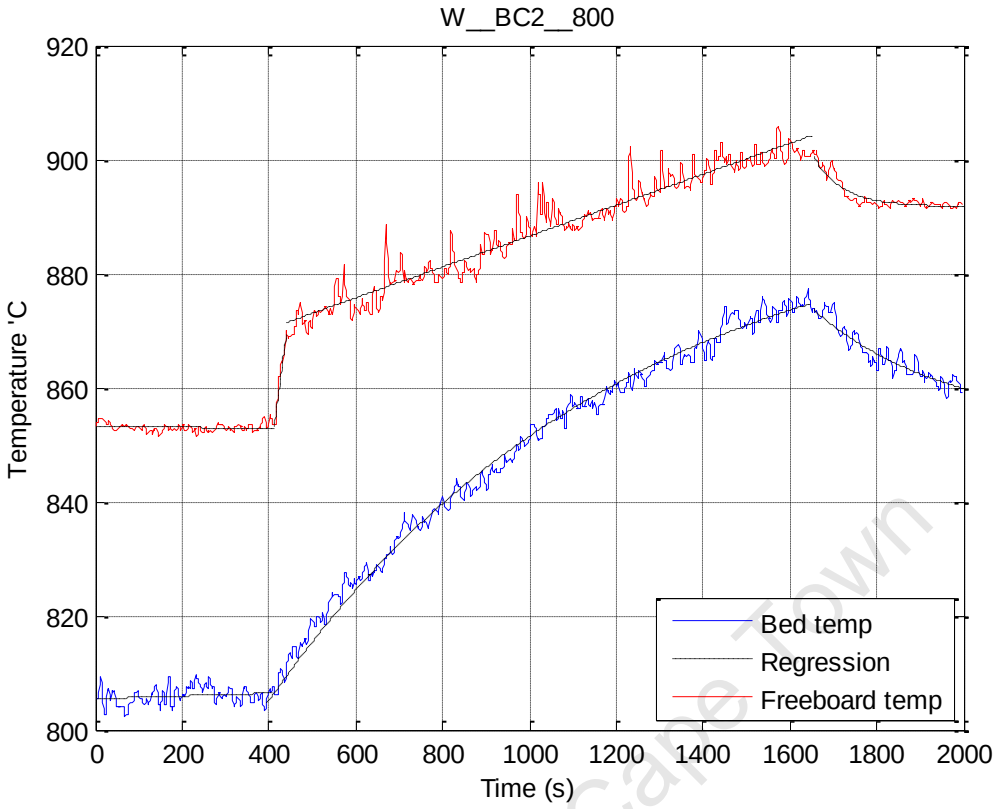
The bed and freeboard temperatures are shown, as well as the regression analysis for the bed temperature. The run ID, as given by Table A-0-1 is given at the top of each plot.

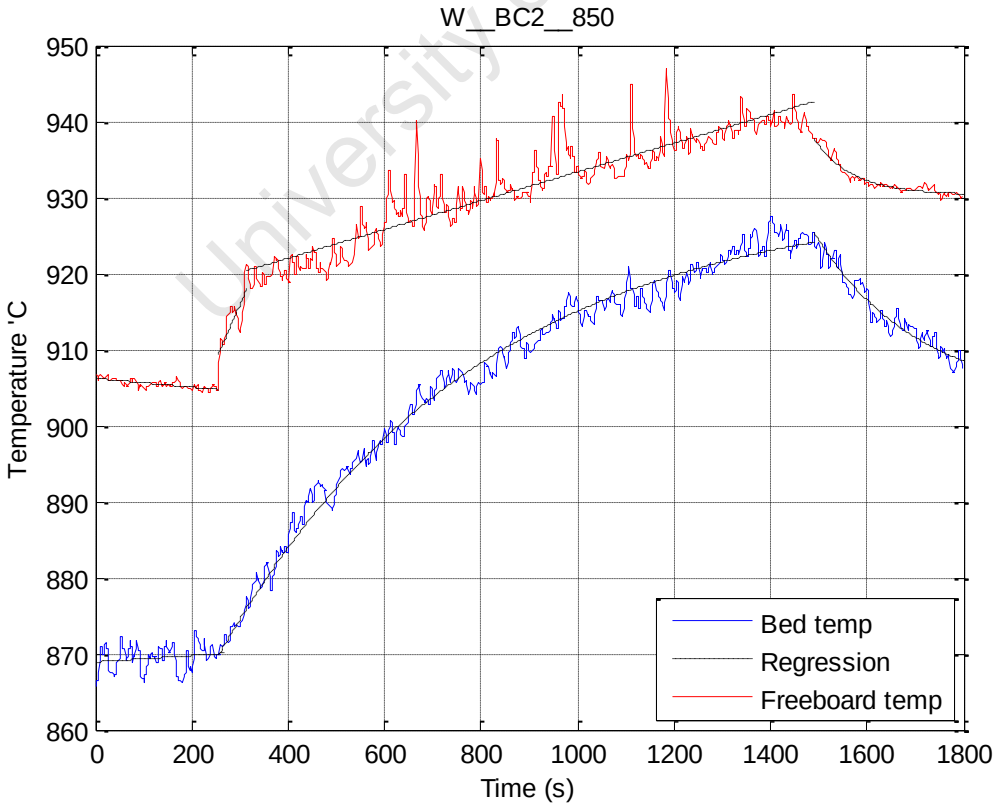


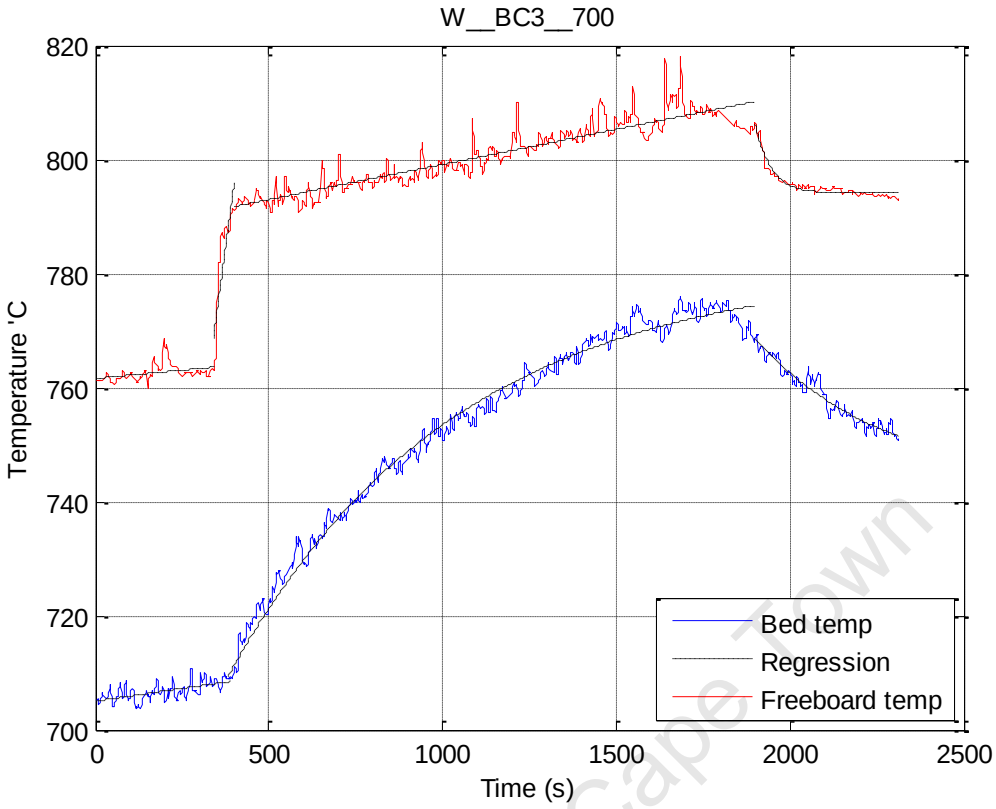


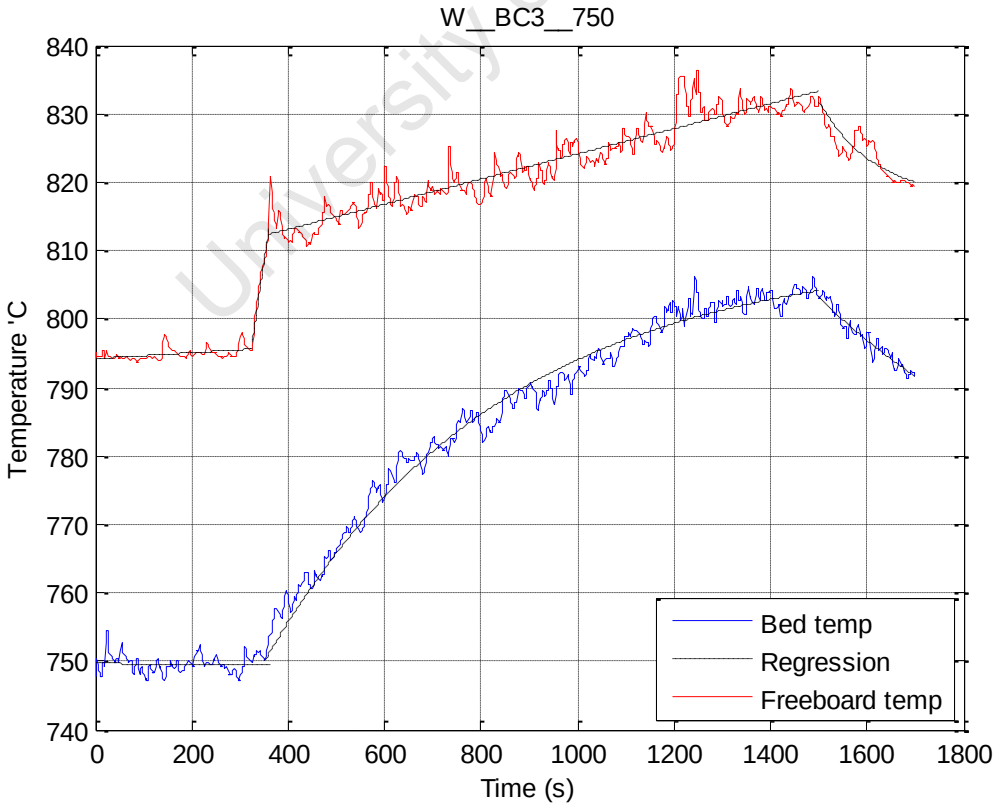


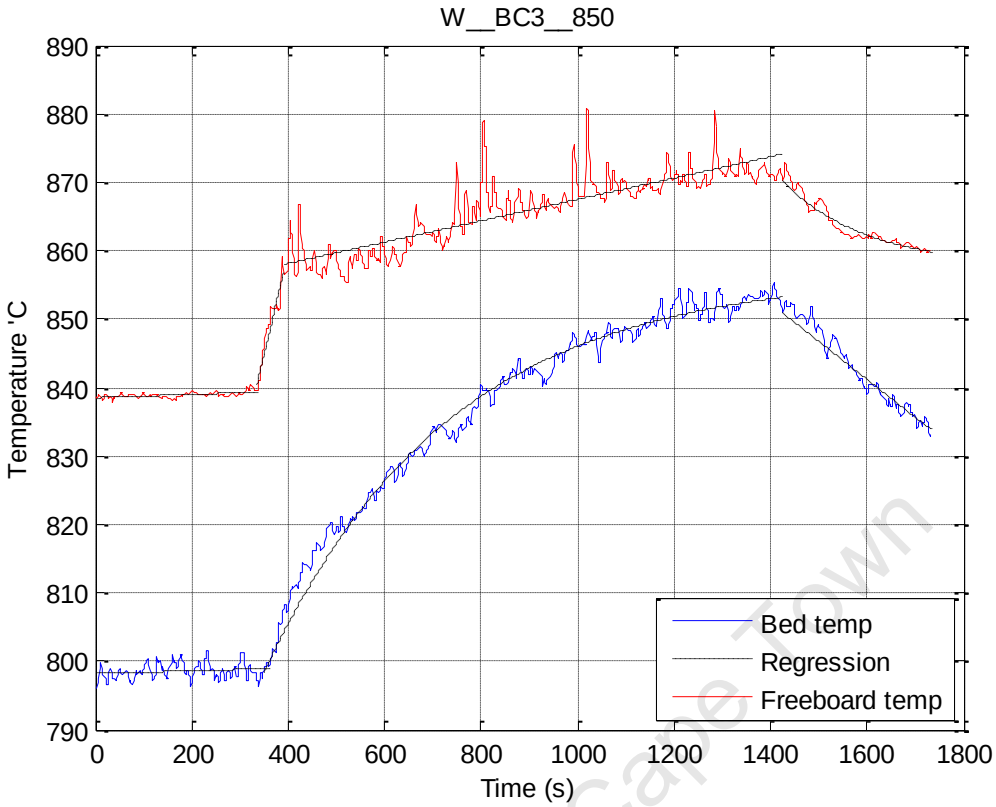


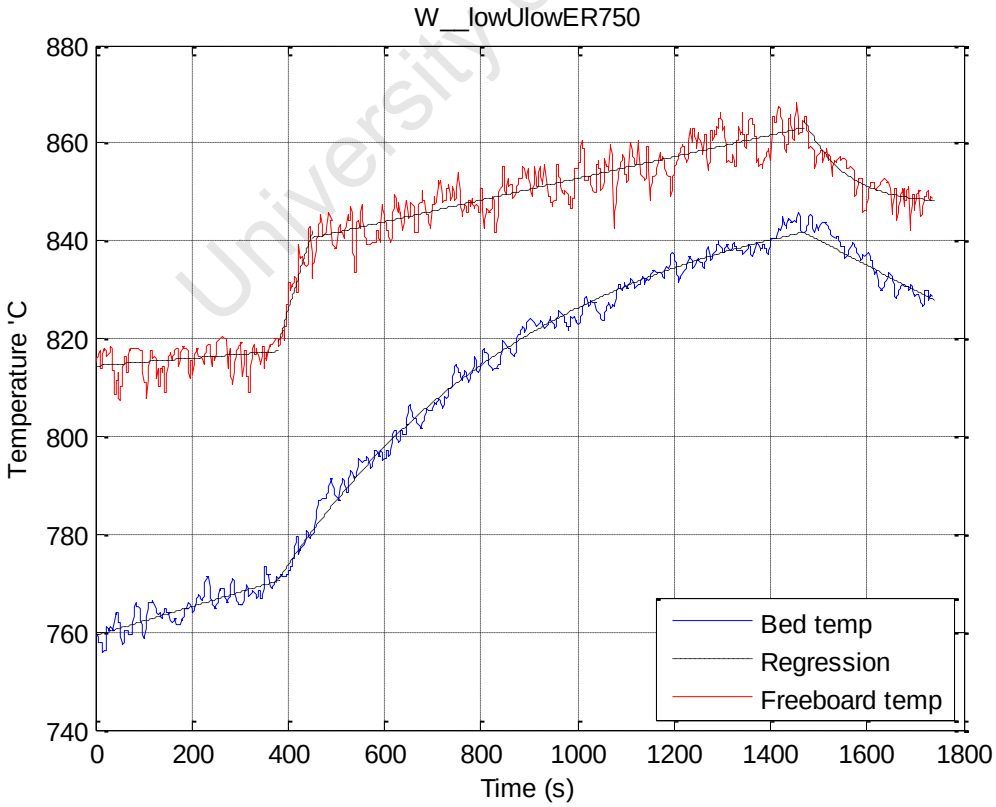


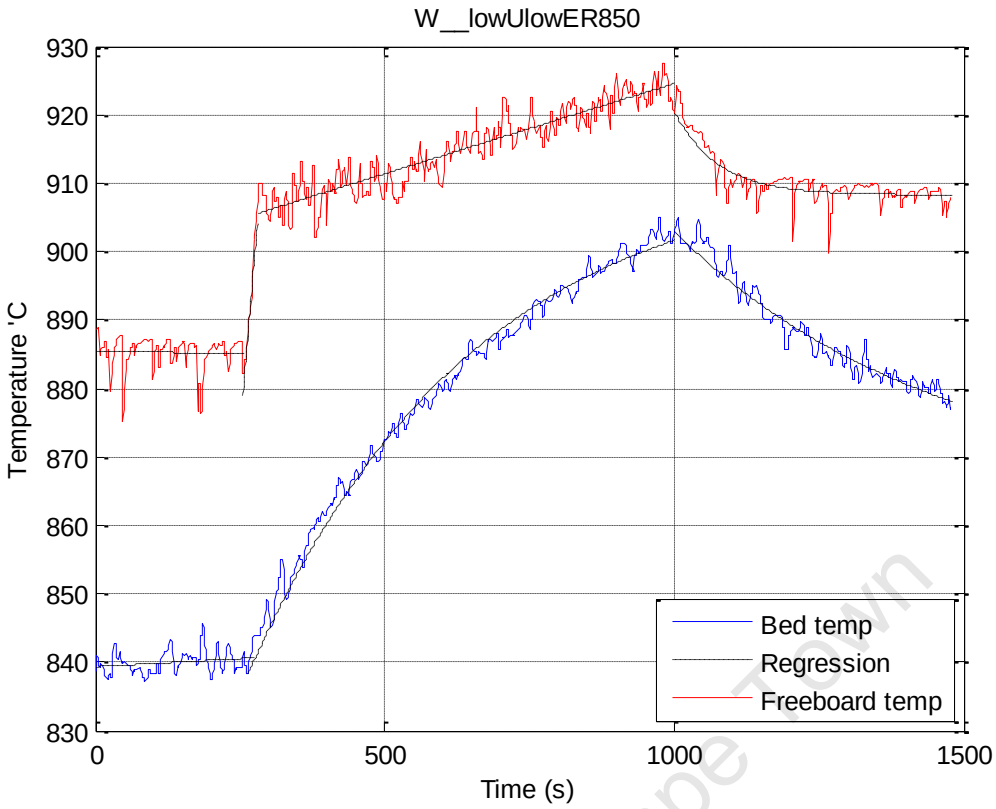


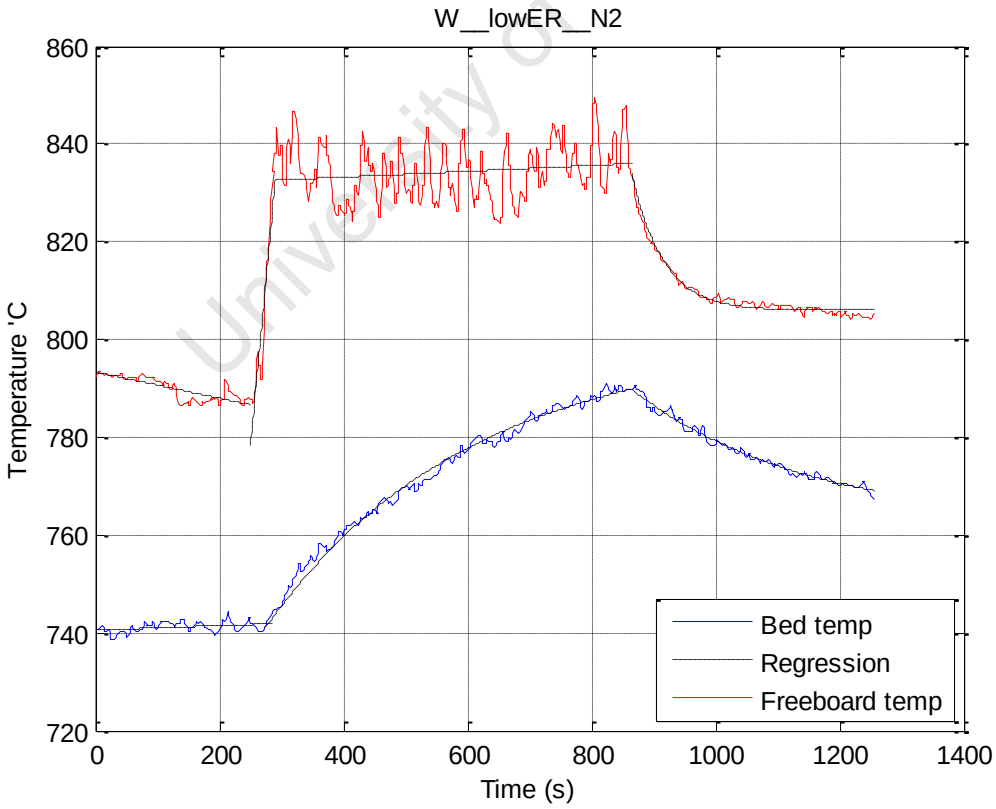


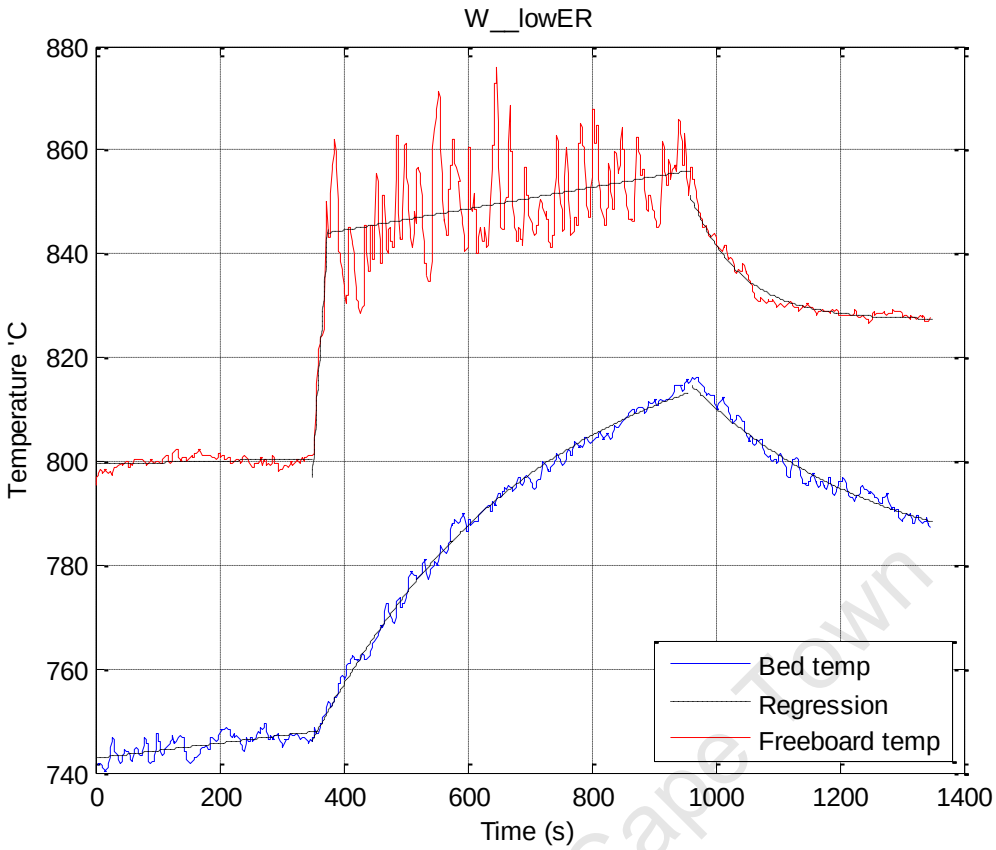


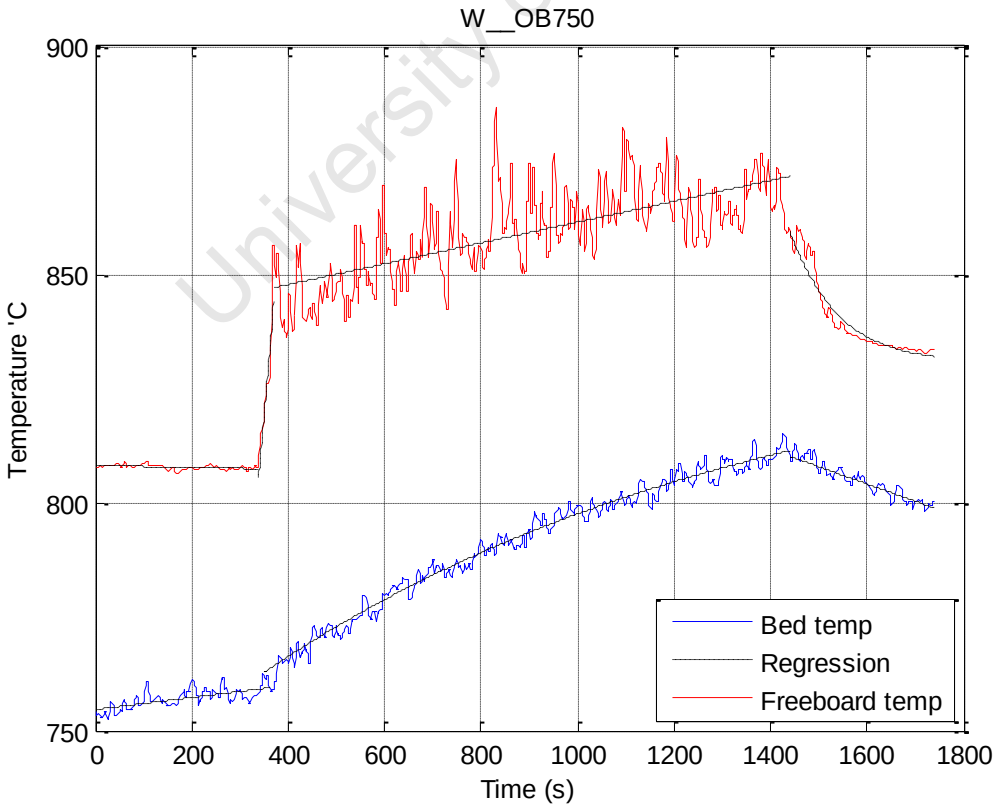


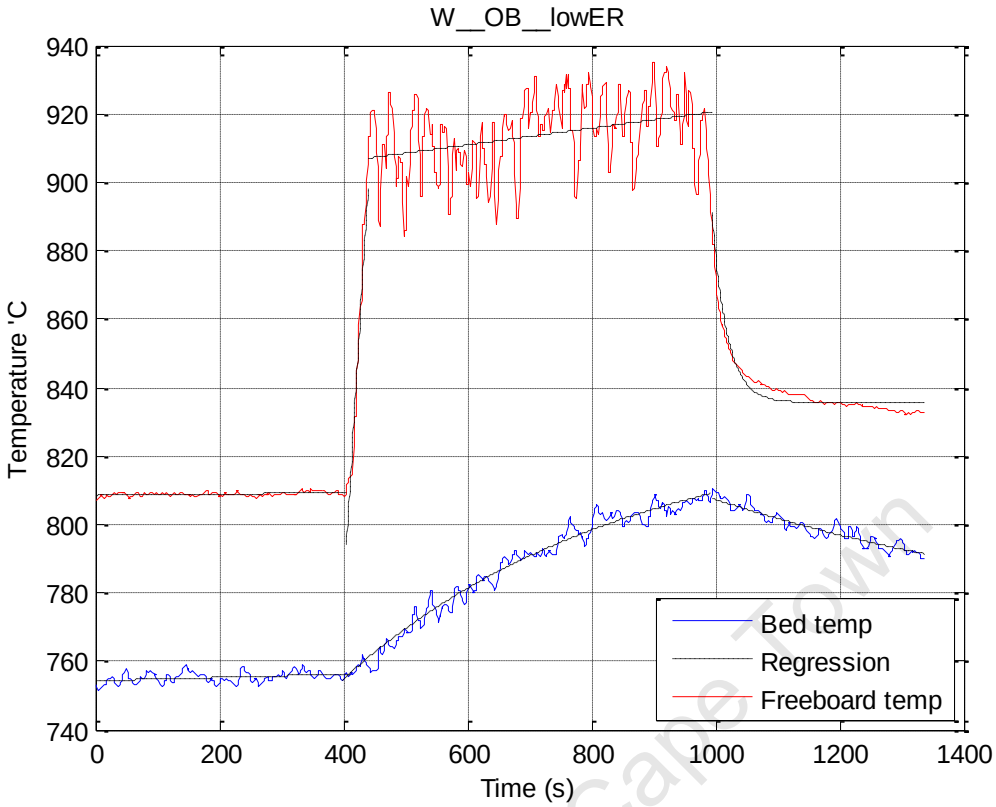


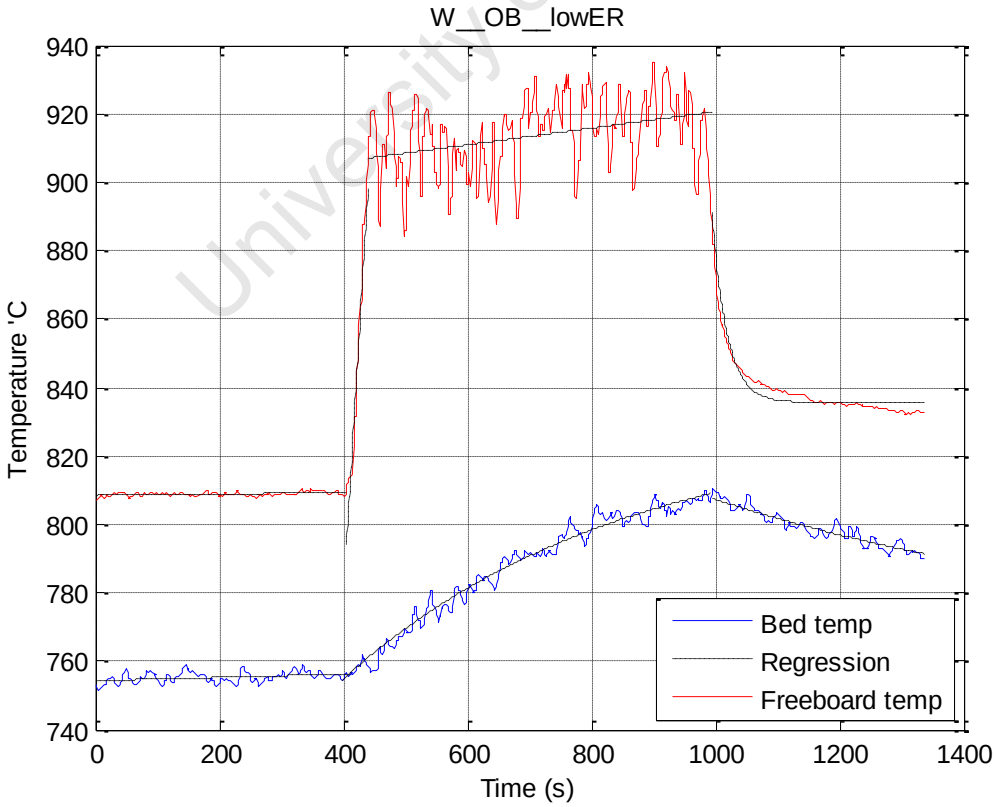


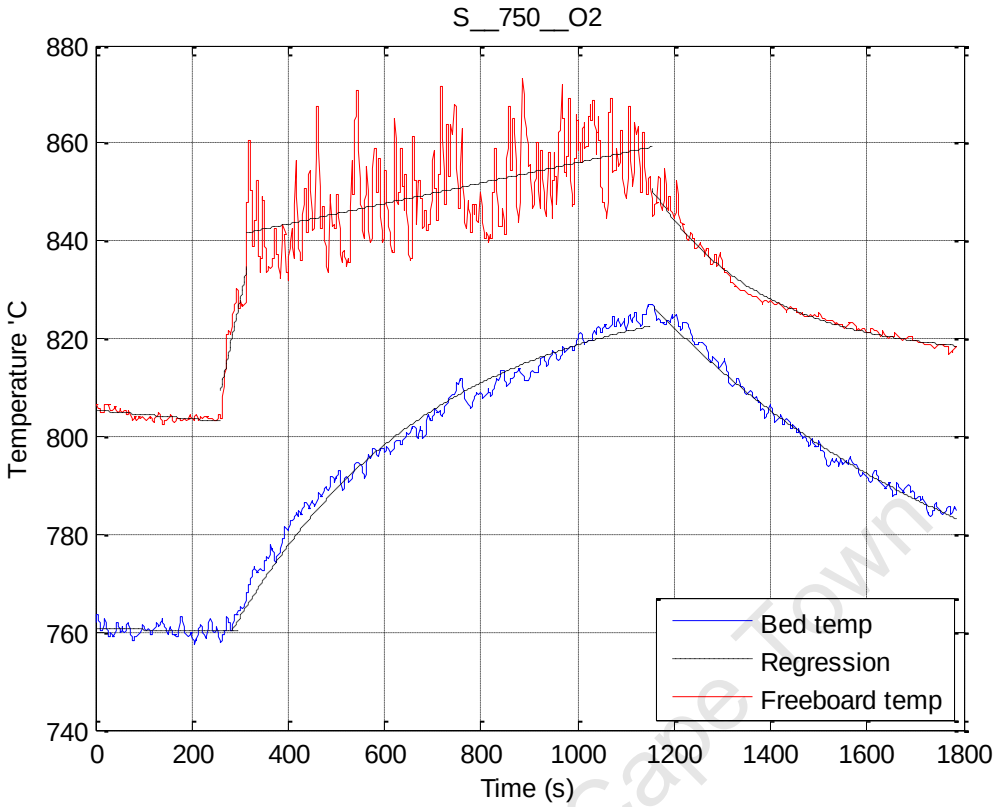


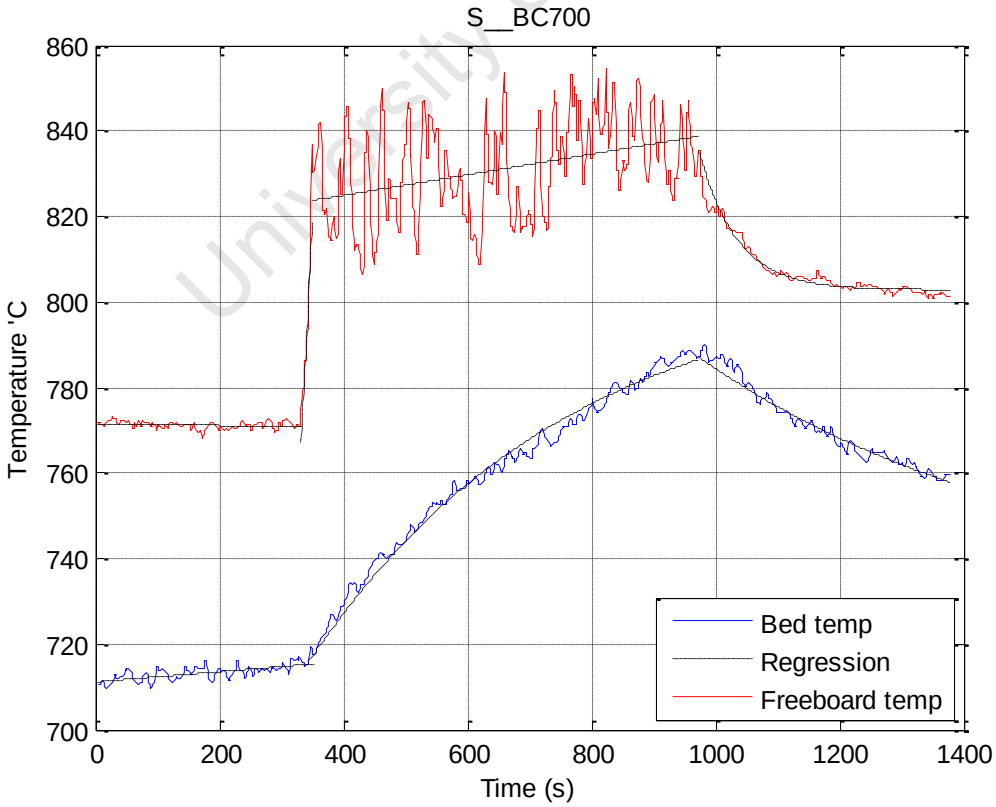


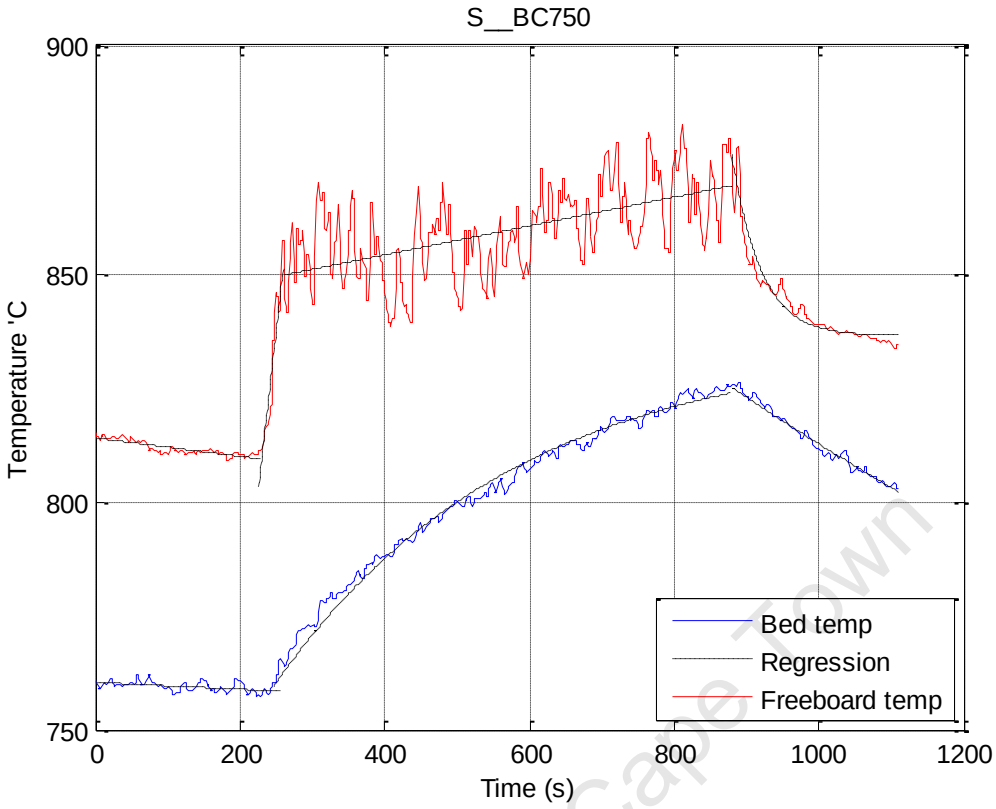


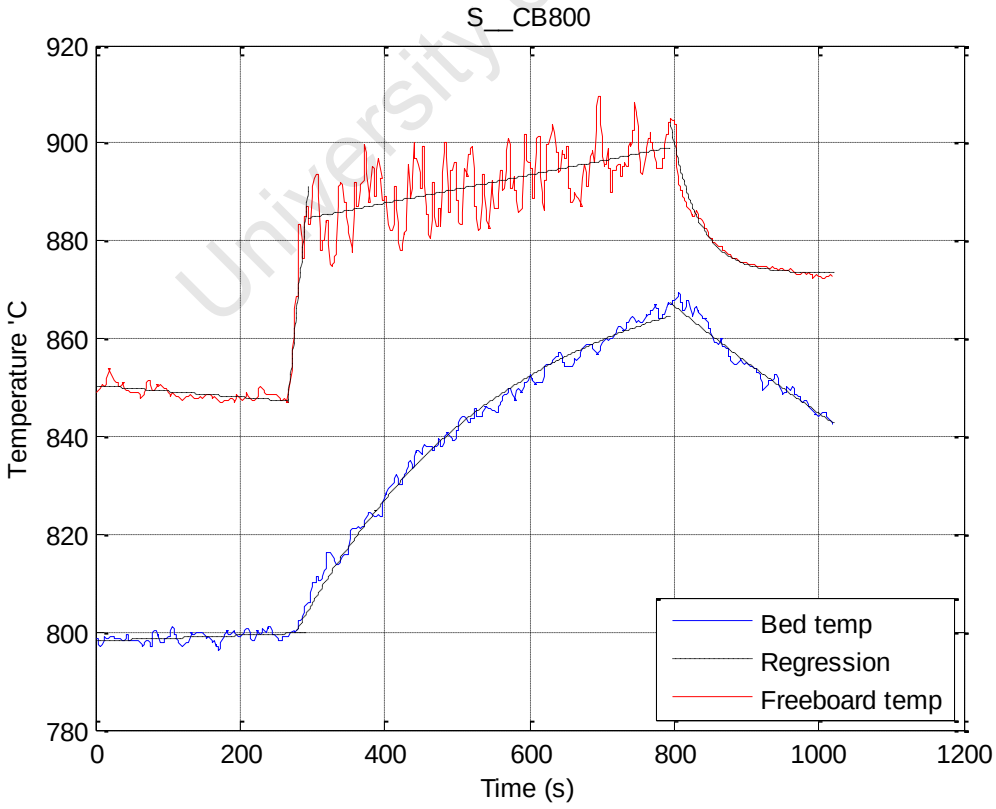


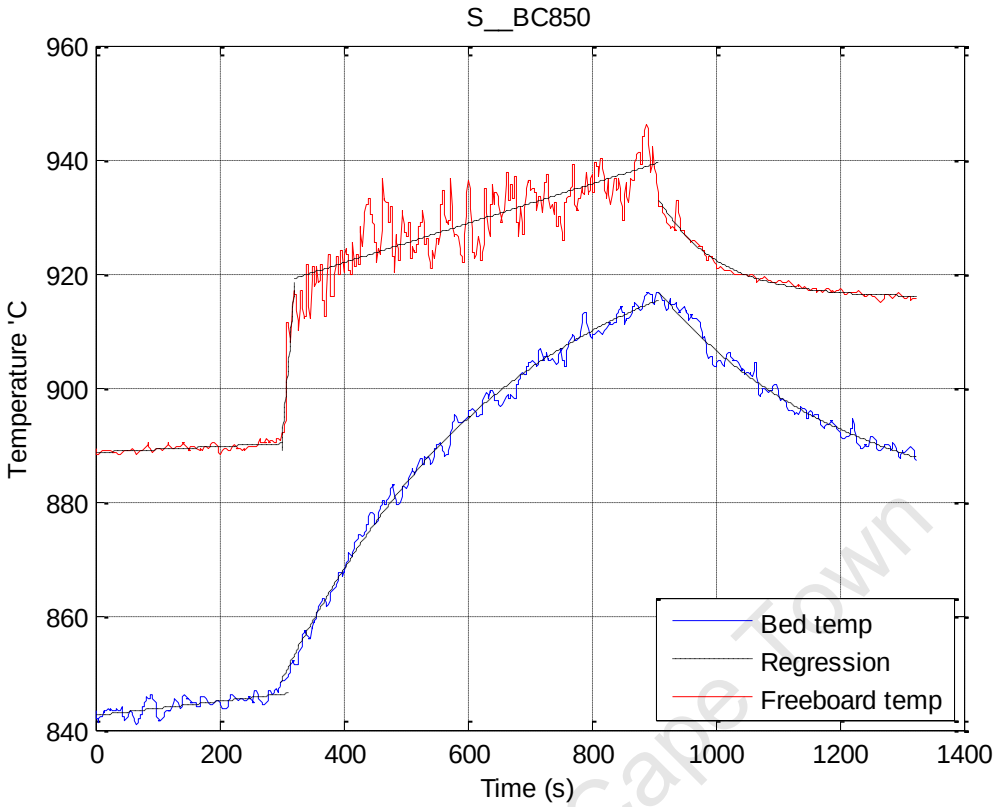


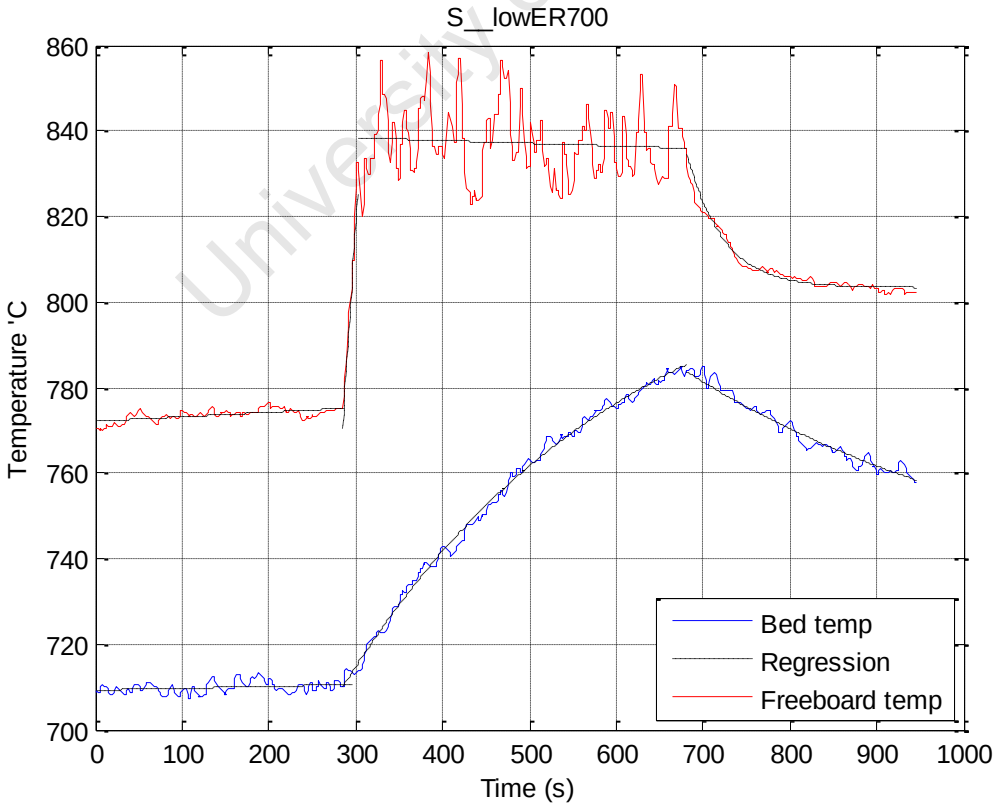












A.4. Matlab code for mass balance

```
% DWLSAR002              26 March 2008
% Mass balancing

close all, clear all, clc

load massbal
% "cAll" is NB one for C bal
% column 1: time, column 2: CO,  column 3: CO2 for runs 1 to 18 as on
% 11/12/07 in lab book (and used thru out)

load exceldata2

% t0 and tn used to get average conc over range
t0 = [20 0 50 20 20 50 0 0 0 150 50 50 50 40 50 25 70 20];
tn = [1120 0 1000 1000 1200 1200 0 0 0 1170 770 600 600 1100 1100 570 950
600];

for i = 1:22
    % need molar flow through reactor
    RR = values(i,1);
    RRN2 = values(i,5);
%     v_air = (0.0001*RR.^2 + ...
%         0.0245*RR + 0.0728)*1000; % flow of air in cc/s
    % / 1.5 to account for operating pressure of 1.5barg
    v_air = 0.3124*RR^2 + 23.328*RR + 89.933; % new calibration

    if RRN2 == 0; % flow of second N2 stream (if present)
        v_N2 = 0;
    else
        v_N2 = (0.0001*RR.^2 + ...
            0.0245*RR + 0.0728)*1000;
    end
    v_hopper = 4; % N2 hopper flow always 4l/min
%     n_air = (v_air+v_N2)/1e6*1e5/8.314/(21+273);
%     n_N2 = v_hopper./1000/60*1e5/8.314/(21+273);
%     pN2(i) = n_N2/(n_N2+n_air(i));
    n_gas = (v_air+v_N2)/1e6*1e5/8.314/(21+273) + ...
        v_hopper./1000/60*1e5/8.314/(21+273) %mol

    % convert V to %CO (column:2) or %CO2 (column:3) (time = column 1)
    % from excel:
    % %CO = 0.0222V - 0.0492
    % %CO2 = 0.0296V - 0.0566
    % %O2 = 0.2489x - 0.0012
    cEnd = find(cAll(:,3*i-2));
    time = cAll(cEnd,3*i-2);
    CO = 0.0222*cAll(cEnd,3*i-1) - 0.0492;
    CO2 = 0.0296*cAll(cEnd,3*i) - 0.0566;

    % Plot selected runs
%     if i ~= 2 & i ~=7 & i ~=8 & i ~=9;
%     figure(i)
%     plot(CO,'g'), grid on, hold on
%     plot(CO2,'b'), grid on, hold on
```

```
    % INSTANTEOUS BALANCE
%     cAve_V = mean(cAll(t0(i):tn(i),3*i));
%     cAve = 0.0296*cAve_V - 0.0566; %fractional output
%     mC_out = cAve*n_gas*12/1000; %in kg C/s
%     end

%     % get biomass flow
%     M_hopper = values(i,2);
%      t1 = values(i,3);  % using the CO2 traces to work out m_bio = total
mass/total time
%     t2 = values(i,4);
%     if i <= 16; % if statement for diff compositions
%         cmp = valuesComp(1,1); % wood
%      else
%         cmp = valuesComp(1,2); % sludge
%      end
%     % call to functions
%     mb = m_bio(M_hopper, t1, t2); % in kg/s
%     mC_in = mb*cmp; % kg C/s

    % CUMULATIVE BALANCE
    % take intergrals under molCO and molCO2
    intCO = 0; intCO2 = 0;
    for k = 1:length(cEnd)-1
        a1 = 0.5*(CO(k)+CO(k+1))*(time(k+1)-time(k));
        intCO = intCO+a1;
        a2 = 0.5*(CO2(k)+CO2(k+1))*(time(k+1)-time(k));
        intCO2 = intCO2+a2;
    end
    molCO = intCO.*n_gas;  % mol CO/mol total*s*mol total/s = mol CO
    molCO2 = intCO2.*n_gas; % mol
    molC_out = molCO+molCO2; %mol
    results(i,1) = molCO;
    results(i,2) = molCO2;

    Cout = 12*(molCO+molCO2); %g/mol*mol => g
    results(i,3) = Cout;

    % Need Cin
    if i <= 16; % if statement for diff compositions
%         C = 0.50;
        C = valuesComp(1,1); %wood
    else
        C = valuesComp(1,2); %sludge
    end
    Cin = values(i,2)*C;  %g*%C w/w => g
    results(i,4) = Cin;
    error = (Cin-Cout)/Cin;
    results(i,5) = error;

    molC_in = Cin/12;  % mol

    res2(i,1) = molC_in;  %****
    res2(i,2) = molC_out; %****

    % can do O2 bal on runs 11 to 17
    % from massbal.mat: oAll = NB. column 1 = time, column 2 = O2
    oBal = 0.5*molCO+molCO2; %mol O2
    results(i,6) =oBal;
```

```
    if i >= 11 & i <= 17
        j = i-10;
        oEnd = find(oAll(:,2*j-1));
        time = oAll(oEnd,2*j-1);
        O2 = 0.2489*oAll(oEnd,2*j) - 0.0012;  % mol/m3
%         plot(O2,'r'), grid on, hold on
        O2 = O2 - O2(1); %"normalise"

        %take intergral under curve and make positive
        intO2 = 0;
        for k = 1:length(oEnd)-1
            a3 = 0.5*(O2(k)+O2(k+1))*(time(k+1)-time(k));
            intO2 = intO2+a3;
        end
        intO2 = -intO2;
        molO2 = intO2.*n_gas;  % mol O2
        results(i,7) = molO2;  %****
        errorO2 = (molO2 - oBal)/molO2;
        results(i,8) = errorO2;
    end

end

wi = 1:1:9; si = 17:21; oi = 11:1:17;
figure(1)
plot(wi,res2(wi,1),'o k', wi, res2(wi,2),'x k' ), grid on, hold on
xlabel('Run number (wood runs)')
ylabel('Carbon mol balance (gmol)')
legend('Carbon in', 'Carbon out')

figure(2)
plot(si,res2(si,1),'o k', si, res2(si,2),'x k' ), grid on, hold on
xlabel('Run number (sludge runs)')
ylabel('Carbon mol balance (gmol)')
legend('Carbon in', 'Carbon out')

figure(3)
plot(oi,results(oi,7), '* k', oi, res2(oi,2),'x k', oi, res2(oi,1), 'o k'
), grid on, hold on
xlabel('Run number')
ylabel('Oxygen mol balance (gmol)')
legend('Total oxygen consumed', 'Oxygen out as CO2 and CO', 'Mols carbon
in')
```


A.5. Matlab code for regression analysis

```
% Analysis2 on 22 datasets

clc, close all, clear all
```

```
load exceldata2
load dataADD2 %in which datatab is NB one

t0s = [341 305 360 377 396 258 380 352 350 372 ...
    265 275 350 350 260 400 285 341 245 280 300 286]';
tfs = [1480 1465 1390 1360 1650 1490 1900 1500 1425 1469 ...
    1002 865 959 1441 1389 994 1153 973 880 795 909 682]';
% tfb0 = [330 290 350 370 415 254 340 325 330 380 260 255 350 340 235 405
262 332 225 265 300 285;
%      360 322 378 395 440 315 400 360 390 405 285 293 375 370 288 450 315
350 260 295 322 302]';

tfb0 = [330 292 350 370 412 254 340 325 335 380 255 250 350 339 234 405 260
332 226 266 300 285;
    355 322 378 395 440 315 400 360 390 450 280 290 375 370 288 440 315 350
260 295 322 302]';

% pre-allocate variables
TT = zeros(61,22); % EB
t = 0:1:60;
res = zeros(22,20); % results from first analysis
res2 = zeros(22,3); %results from second analysis - looking at slopes
res3 = zeros(22,3);

% constants
M_bed = 2.15;  %kg
Cp_bed = 870;  % J/K/kg (ref: www.azom.com/details.asp?ArticleID=52 ... for
94% pure Al2O3 (ours was 95.2%))
bedHeight = 0.35; % m (manometer reading)
wallThickness = 0.005; % m
pSteel          =          8000;       %          kg/m3          (ref:
www.azom.com/details.asp?ArticleID=863#_Physical_Properties)
Di = 0.078; % m
M_s = (bedHeight*pi*Di*wallThickness)*pSteel;


for i = 1: 22;
    t0 = t0s(i);
    tf = tfs(i);
    Y = datatab(find(datatab(:,2*i-1)),2*i-1);
    Tb0 = Y(t0); %bed temp as feeding commences
    Tbf = Y(tf); %bed temp as feeding ends
    tn = length(Y);
    % find slope of first portion (ss1)for BED
    m1 = polyfit([0:1:t0-1]',Y([1:1:t0]'),1);
    ss1 = polyval(m1,[0:1:t0+10]');
    Tss1 = mean(Y([1:1:t0]'));
    COVss1 = std(Y([1:1:t0]'))/mean(Y([1:1:t0]'))*100;

    % find slope of second portion (temp increase)
    m2 = polyfit((t0:1:t0+45)', Y(t0:1:t0+45,1),1);
    trans = polyval(m2,(t0-10:1:t0+45)');

    % BED ANALYSIS
    % inputs for minimisation routines
    xdat = (0:1:tf-t0)'; %for nlinfit
    ydat = Y(t0:1:tf) - Y(t0);  % in deviation variables
    % nlinfit to get CI on estimates
```

```
    [beta,resid,J] = nlinfit(xdat,ydat,@expnlinfit,[50 0.0001]);
    NLF = beta(1)*(1-exp(-beta(2).*xdat))+Tb0;
    % get CI on estimates
    ci = nlparci(beta,resid,J);
    dTdt0 = beta(1)*beta(2); % initial rate from fitted curve (nlinfit)
    dTdt0_LCI = ci(1,1)*ci(2,1);
    dTdt0_UCI = ci(1,2)*ci(2,2);

    % FREEBOARD ANALYSIS
    % find slope of ss1
    Y2 = datatab(find(datatab(:,2*i)),2*i);
    TfbF = Y2(tf);
    mfb1 = polyfit([0:1:tfb0(i,1)-1]',Y2(1:1:tfb0(i,1)'),1);
    Tfb1_fit = polyval(mfb1,[0:1:tfb0(i,1)]');
    Tfbss1 = mean(Y2(1:1:tfb0(i,1))');

    % find rate of increase in fb temp
    Tfb0 = Y2(tfb0(i,1)); % start of fb temp rise
    Tfb2 = Y2(tfb0(i,2)); % end fb temp rise
    mfb2 = polyfit(tfb0(i,1):1:tfb0(i,2)',Y2(tfb0(i,1):1:tfb0(i,2))',1);
    Tfb2_fit = polyval(mfb2, tfb0(i,1):1:tfb0(i,2)');

    % find slope of freeboard during feeding
    % (Over entire period of feeding)
    mfb3 = polyfit((tfb0(i,2):1:tf)', Y2(tfb0(i,2):1:tf'),1);
    Tfb3_fit = polyval(mfb3,(tfb0(i,2):1:tf)');
    dT_fb = Tfb3_fit(1) - Tfb0;



    % END OF RUN ANALYSIS
    xdatN = [0:1:tn-tf]';
    ydatN = Y(tf:1:tn)-Tbf;
    [betaN] = nlinfit(xdatN,ydatN,@expnlinfit2,[50 0.02]);
    NLFN = -betaN(1)*(1-exp(-betaN(2).*xdatN))+Tbf;
    ydatNfb = Y2(tf:1:tn)-TfbF;
    [betaNfb] = nlinfit(xdatN,ydatNfb,@expnlinfit2,[50 0.02]);
    NLFNfb = -betaNfb(1)*(1-exp(-betaNfb(2).*xdatN))+TfbF;
    mN(i) = betaN(1)*betaN(2);
    mNfb(i) = betaNfb(1)*betaNfb(2);

    % dTfb_b analysis
    dTfb_b0 = Y2 - Y;
    betaD1 = nlinfit((tfb0(i,2):1:tf)',...
        dTfb_b0(tfb0(i,2):1:tf)-dTfb_b0(tfb0(i,2)),@expnlinfit2,[20,0.002])
    dTfb_b1                          =                          -betaD1(1)*(1-exp(-
betaD1(2).*(tfb0(i,2):1:tf)'))+dTfb_b0(tfb0(i,2));

    % EB MODEL
    % exceldata for EB calc
    RR = values(i,1);
    M_hopper = values(i,2);
    t1 = values(i,3);   % using the CO2 traces to work out m_bio = total
mass/total time
    t2 = values(i,4);
    RRN2 = values(i,5);
    if i <= 16 % if statement for diff compositions
        comp = valuesComp(:,1);
        LHVar = 14e6;
        %disp('wood')
```

```
    else
        comp = valuesComp(:,2);
        LHVar = 19e6;
        %disp('sludge')
    end
    % call to functions
    mb = m_bio(M_hopper, t1, t2);
    [mg U Umf ER LHV] = EBinputs(RR, RRN2, Tb0,comp,mb);
    mg_i(i) = mg; %for calculating hA and Te
    Cpa = Cp_air(Tb0+273);
    Cpa_i(i) = Cpa;
    Cps = CpS(Tb0+273);

    % Calc Q,fb
    Cpa_fb(i) = Cp_air(Tfb3_fit(1));
    UAfb = 0.4249; %from regression analysis
    Qc_fb(i) = (mg*Cpa_fb(i))*dT_fb;

    %from report: A = (mbiomass*Hc)/(mair*Cpair)    [1000 K]
                 %alpha  = (mair*Cpair)/(Mbed*Cpbed)  [0.0002 s-1]
%     UA = 0.4468; %from regression analysis of Tele vs Tbed
    alpha = (mg*Cpa)/(M_bed*Cp_bed+M_s*Cps);
%     alpha = (mg*Cpa)/(M_bed*Cp_bed);
    A = (mb.*LHVar)/(mg*Cpa);



    % EB model
    for j = 1:length(t)
        TT(j,i) = Tb0 + A*(1-exp(-alpha*t(j)));
    end

    % RESULTS TABLE
    % parameters
    res(i,1) = Tb0; % initial bed temp
    res(i,2) = mb*1000*60;  %report in g/min
    res(i,3) = U/Umf;
    res(i,4) = ER;
    res(i,5) = RR;
    % EB model results
    res(i,6) = dTdt0; % initial slope according to regression
    res(i,7) = m1(1)/(mb*1000*60); % initial slope: linear fit
    res(i,8) = A*alpha; % **** initial slope according to EB (normalised to
feedrate)
    % freeboard results
    res(i,9) = Tfb0; %initial fb temp
    res(i,10) = mfb3(1); % slope of fb during feeding
    res(i,11) = dT_fb; %jump in fb temp
    res(i,12) = Tfb0 - Tb0; % initial diff betw bed and fb temp
    %     res(i,13) = Tfb2_fit(1) - Tb2; % final diff betw bed and fb temp
    res(i,14) = mfb2(1); % slope of fb temp jump

    % compare regression coeffs and calc'd coeffs
    res(i,15) = A; % from EB
    res(i,16) = alpha; % from EB
    res(i,17) = beta(1); % fitted data from nlinfit
    res(i,18) = beta(2); % fitted data from nlinfit
    res(i,19) = dTdt0_LCI/(mb*1000*60);
    res(i,20) = dTdt0_UCI/(mb*1000*60);
    res2(i,1) = dTdt0/(A*alpha); %dTdt0 = beta(1)*beta(2)
    res2(i,2) = dTdt0_LCI/(A*alpha);
```

```
    res2(i,3) = dTdt0_UCI/(A*alpha);

    % Results display: latest
    % In absolute terms
    res3(i,1) = mb.*LHVar; %max heat released
    res3(i,2) = (M_bed*Cp_bed+M_s*Cps)*dTdt0; %actual heat released to bed
    res3(i,3) = Qc_fb(i); %heat released as overbed burning
    % As fractions
    res3(i,4) = (M_bed*Cp_bed+M_s*Cps)*dTdt0/(mb.*LHVar);
    res3(i,5) = (M_bed*Cp_bed+M_s*Cps)*dTdt0_LCI/(mb.*LHVar);
    res3(i,6) = (M_bed*Cp_bed+M_s*Cps)*dTdt0_UCI/(mb.*LHVar);
    res3(i,7) = Qc_fb(i)/(mb.*LHVar);

    % PLOTS
%     figure(i)
% %     subplot(2,1,1)
%     h(1) = plot(0:1:tn-1,Y(1:1:tn),'b'),grid on, hold on;
%     plot(ss1,'k --'), grid on, hold on
%     %     plot((t0-10:1:t0+45)',trans, 'b'), grid on, hold on
%     h(2) = plot(t+t0,TT(:,i), 'k', 'LineWidth',2 ), grid on, hold on;
%     %     plot(xdat+t0-1,FittedCurve+Tb0, 'k'),grid on, hold on
%     h(3) = plot(xdat+t0-1,NLF, 'k --'),grid on, hold on; % nlinfit
%     h(4) = plot((0:1:tn-1)', Y2(1:1:tn)', 'r'), grid on, hold on;
%     plot((tfb0(i,2):1:tf)', Tfb3_fit, 'k --'), grid on, hold on
%     plot(Tfb1_fit,'k --'), grid on, hold on
%     plot(tfb0(i,1):1:tfb0(i,2)',Tfb2_fit, 'k --'), grid on, hold on
%     plot((tf:1:tn)', NLFN, 'k --'), grid on, hold on
%     plot((tf:1:tn)', NLFNfb, 'k --'), grid on, hold on
%     xlabel('time (s)')
%     ylabel('Temperature ''C')
%     title(runInfo(i+1))
%                     legend(h,'Bed    temp','Model','Regression','Freeboard
temp','Location','SE')

%     subplot(2,1,2)
%     plot(dTfb_b0), grid on, hold on
%     plot((tfb0(i,2):1:tf)',dTfb_b1, '-- k'), grid on, hold on
%     plot(tfb0(i,2), dTfb_b0(tfb0(i,2)), 'o'), grid on , hold on
%     xlabel('Time (s)')
%     ylabel('Difference in temperature between bed and freeboard (''C)')

end


figure(23) % wood runs percent heat released
h(1) = plot(res([1 2 3 4 5 6 8 ],1),res3([1 2 3 4 5 6 8 ],4),'o k'), hold
on
h(2) = plot(res([1 2 3 4 5 6 8 ],1),res3([1 2 3 4 5 6 8 ],5),'+ k'), hold
on
plot(res([1 2 3 4 5 6 8 ],1),res3([1 2 3 4 5 6 8 ],6),'+ k'), hold on, grid
on
h(3) = plot(res([1 2 3 4 5 6 8 ],1),res3([1 2 3 4 5 6 8 ],7),'* k'), hold
on
xlabel('Initial bed temperature (''C)');
ylabel('Fraction of heat released');
legend(h,'Fraction of heat released to bed',...
    '95% confidence interval from regression estimates',...
    'Fraction of heat to freeboard')
```

```
figure(24) % sewerage sludge percent heat released
h(1) = plot(res([17:21],1),res3([17:21],4),'o k'), hold on
h(2) = plot(res([17:21],1),res3([17:21],5),'+ k'), hold on
plot(res([17:21],1),res3([17:21],6),'+ k'), hold on, grid on
h(3) = plot(res([17:21],1),res3([17:21],7),'* k'), hold on
xlabel('Initial bed temperature (''C)');
ylabel('Fraction of heat released');
legend(h,'Fraction of heat released to bed',...
    '95% confidence interval from regression estimates',...
    'Fraction of heat to freeboard')

% Overbed bed runs heat released
figure(25)
h(1) = plot(res([14:15],1),res3([14:15],4),'o k'), hold on
h(2) = plot(res([14:15],1),res3([14:15],5),'+ k'), hold on
plot(res([14:15],1),res3([14:15],6),'+ k'), hold on, grid on
h(3) = plot(res([14:15],1),res3([14:15],7),'* k'), hold on
xlabel('Initial bed temperature (''C)');
ylabel('Fraction of heat released');
legend(h,'Fraction of heat released to bed',...
    '95% confidence interval from regression estimates',...
    'Fraction of heat to freeboard')

figure(26) % wood runs fb temp jump
plot(res([1:1:8],1),res([1:1:8],11),'o k'), hold on, grid on
xlabel('Initial bed temperature (''C)');
ylabel('Jump in freeboard temperature (''C)');
title('Freeboard temperature jump for wood runs')

figure(27) %sewerage sludge fb temp jump
plot(res([17:21],1),res([17:21],11),'o k'), hold on, grid on
xlabel('Initial bed temperature (''C)');
ylabel('Jump in freeboard temperature (''C)');
title('Freeboard temperature jump for sludge runs')



% Overbed fb temp jump
figure(28)
plot(res([14:15],1),res([14:15],11),'o k'), hold on, grid on
xlabel('Initial bed temperature (''C)');
ylabel('Jump in freeboard temperature (''C)');
title('Freeboard temperature jump for overbed wood runs')

% Initial slope at end of run
figure(29)
plot(mN,mNfb, 'o k'), grid on, hold on
xlabel('Initial rate of cooling for bed (K/s)')
ylabel('Initial rate of cooling for freeboard (K/s)')

% represent as bar chart
figure(30)
bb(1) = bar([1:8],res3(1:8,1),'w')
hold on
bb([2:3])= bar([res3(1:8,2) res3(1:1:8,3)],'stack')
colormap([0 1 1;0 0 0]);
xlabel('Run number')
ylabel('Energy flow (W)')
legend(bb, 'Maximum heat released to bed',...
```

```
'Actual heat released to bed','Heat released by overbed
burning','location','SE')


figure(31)
bb(1) = bar([17:21],res3([17:21],1),'w')
hold on
bb([2:3])= bar([17:21],[res3([17:21],2) res3([17:21],3)],'stack')
colormap([0 1 1;0 0 0]);
xlabel('Run number')
ylabel('Energy flow (W)')
legend(bb, 'Maximum heat released to bed',...
'Actual heat released to bed','Heat released by overbed
burning','location','SE')

figure(32)
bb(1) = bar([14:15],res3(14:15,1),'w')
hold on
bb([2:3])= bar([14:15],[res3(14:15,2) res3(14:15,3)],'stack')
colormap([0 1 1;0 0 0]);
xlabel('Run number')
ylabel('Energy flow (W)')
legend(bb, 'Maximum heat released to bed',...
'Actual heat released to bed','Heat released by overbed
burning','location','SE')

tableS = cell(23,5);
tableS(1,1) = {'Run number'};
tableS(2:23,1) = num2cell([1:1:22]');
tableS(1,2) = {'Run ID'};
tableS(2:23,2) = runInfo(2:23,1);
tableS(1,3:6) = {'Initial bed temperature (''C)','Fuel flowrate
(g/min)',...
    'U/Umf','ER'};
tableS(2:23,3:6) = num2cell(res(:,[1:4]));

% Find UA for bed
ii = [1:1:9 12:1:22]; %select conditions of same gas flowrate
I = 22/2; % amps
V = 240/3; %volt
sigma = 5.67e-8; %W/m2/K^4;
eta = 0.7; % pers comm. Kanthal
Aele = pi*0.012*0.20;
Tele = ((values(ii,6)/100)*I*V/sigma/eta/Aele).^(1/4); % re-arrange Q =
sigma*eta*T,blackbody^4
w2 = polyfit(res(ii,1),Tele,1);
aveCpa = mean(Cpa_i(ii));
aveMg = mean(mg_i(ii));
UA = aveMg*aveCpa/(w2(1)+1);
UA2 = -aveMg*aveCpa*294/w2(2);
U = UA/(Di*pi*0.35);

% Find UA for freeboard
% Use mean temp diff of freeboard
Tfb_ave = (res(ii,9)-res(ii,1))/2+res(ii,1);
w2fb = polyfit(res(ii,9), Tele,1);
aveCpa_fb = mean(Cpa_fb(ii));
UAfb = aveMg*aveCpa_fb/(w2fb(1)+1);
Ufb = UAfb/pi/Di/0.46;
figure(33)
```

```
plot(res(ii,1),Tele-273, '* k'), grid on, hold on
plot(res(ii,1), polyval(w2,res(ii,1))-273,'k'), grid on, hold on
plot(res(ii,1),res(ii,1),'-+'), grid on, hold on
plot(Tfb_ave, Tele-273, 'o b'), grid on, hold on
% plot(Tfb_ave, polyval(w2fb,res(ii,9))-273, 'b'), grid on, hold on
plot(Tfb_ave, Tfb_ave, '-+ r'), grid on, hold on
xlabel('Initial bed temperatur ''C')
ylabel('Element temperature ''C')

% Plot SS2 presdicted temps
% figure(30)
% plot(res(:,1), res(:,15), 'o r'), grid on, hold on
% plot(res(:,1), res(:,17), 'o k'), grid on, hold on
% legend('Tss2 from EB','Tss2 from regression')
% ylabel('Predicted Tss2 ''C')
% xlabel('Initial bed temp')
```

Appendix B Waste availability calculations

B.1. Derivation of biogas yield from COD destruction

Biogas yield from volatile solids destruction may be derived from first principles in the following manner. The COD of methane is the amount of O_2 required to completely oxidise 1 mol of methane. From the stoichiometry, it can be seen two mols of O_2 are required. Two mols of O_2 are equal to 64 g. Hence the COD equivalent of one mol of methane is 64g COD/mol (Tchobanoglous *et al.*, 2003).

At STP (0°C, 1 atm), 1 mol of ideal gas occupies 22.4 l, hence

22.4 l/mol CH_4 ÷ 64 gCOD/mol CH_4 = 0.35 l CH_4/g COD = 0.35m^3 CH_4/kg COD

B.2. Energy yields from DME White Paper on renewable energy

As the data in the DME paper was quoted in terms of ranges of biomass and energy yields, some further calculations were necessary to obtain a single meaningful value. To relate the biomass energy yield to the areal energy yield a single representative yield was calculated from data from the South African Crop Estimates Committee (CEC) as show in Table B-0-7 The weighted average yield was calculated to be 2.7 ton/Ha Assuming that the lowest biomass yield referred to the lowest areal energy yield, linear interpolation was used to calculate an energy yield corresponding to the weighted biomass yield. This gave an areal energy yield of 40.2 GJ/Ha/yr as shown in

Table B-0-8. Other manipulations are shown in Table B-0-9.

Table B-0-7: Average South African agriculatural seed yield

	Maize	**Sorghum**	**Wheat**	**Sunflower**
Season	06/07 season	06/07 season	Winter 07	06/07 season
Area (Ha)	2,551,800	69,000	632,000	316,350
Crop (ton)	7,125,000	176,000	1,714,950	300,000
Yield	2.8	2.6	2.7	0.9
Yield of residues assuming HI = 0.5	2.8	2.6	2.7	0.9

Table B-0-8: Interpolation for areal energy yield

	Lower limit	Upper limit	Interpolated point
Biomass yield range (ton/Ha/yr)	0.2	10	2.7
Energy yield range (GJ/Ha/yr)	2	140	40.2

Table B-0-9: Manipulations of DME data

	Quoted in White Paper		Conversion to PJ/yr
Bagasse	47.6	PJ/yr	47.6
Paper and sawmills	12167	GWh/yr	43.8
Agricultural	40.2	GJ/Ha/yr	143.5 [a]
Grass	84	GJ/Ha/yr	25.2 [b]
Feedlot manure	5612	GWh/yr	20.2

a.) Based on 3,569,150 Ha planted (2006/2007 season) (Crop Estimates Committee, 2008)
b.) The DME (2004) reports an energy yield of 84GJ/Ha/yr is possible in the low-lying areas of KZN, Eastern Cape and Mpumalanga. Thus the area used in the calculation was based on an area the size of the 2007 sugar cane crop (SASA, 2008)

B.3. Landfill gas electricity generating potential calculations

The ratio of CER's for flaring and electricity generation is calculated as follows:

Flaring: 1 kg CH_4 has GWP of 21 kg CO_2eq

Electricity generation: $E_{thermal}$ = 1 kg CH_4 · 0.0625 kmol/kg · 8.026e8 J/kmol

$E_{electrical} = E_{thermal}$ · conversion efficiency of turbine = 3.5 kWhr

As the electricity is replacing that of Eskom's coal fired power stations with an emission factor of 0.98 kg CO_2/kWhr, the amount mitigated is 3.4 kg CO_2 per kg CH_4.

Data from the Project Design Document cited manufacturer's conversion efficiencies of 26%.